

**TECHNICAL AND ECONOMIC ANALYSIS OF AN ENZYMATIC HYDROLYSIS  
BASED ETHANOL PLANT**

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## **Preface**

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## Section 1.0

### Summary

An analysis of the technical and economic status of the biomass-to-ethanol process was conducted for the Ethanol Program of the U.S. Department of Energy's (DOE) Biofuels System Division. The purpose was to redesign and update the process technology since the last time an analysis was performed (Wright 1988). Since that time, the process and economic parameters have been changed and redefined, significantly changing and improving the technology. The primary objective of this work was to establish goals and direction for future research for the production of ethanol from lignocellulosic biomass.

An economic analysis is performed on a fuel ethanol (90% ethanol, 5% water, and 5% gasoline) plant producing approximately 58 MM gal/yr. The feedstock to the plant is assumed to be whole-wood tree chips delivered to the site for \$42/dry ton. The chips are stored in piles and then delivered on a first-in, first-out basis to a disk refiner for milling to 2.0-mm to 3.0-mm particles. The milled particles are pretreated with dilute acid at 160°C for 10 min. After flash cooling, the slurry is neutralized with lime and a small side stream is pumped to the cellulase production fermenters, while the rest of the stream is pumped to the xylose fermenters. Xylose fermentation is performed by a genetically engineered *Escherichia coli* continuously in a series of fermenters. Cellulase is produced by *Trichoderma reesei* in three batch fermenters. The cellulase is combined with the stream out of the last xylose fermenter, yeast inoculum is added, and simultaneous saccharification and fermentation (SSF) is performed continuously in another series of fermenters. The dilute beer stream from the last SSF fermenter is sent to ethanol purification for concentration of the ethanol to 95 wt %. Then, 5% gasoline is added to denature the fuel. The stream from the bottom of the beer column is sent to centrifugation to remove the solids, which are then burned in a boiler to produce steam and electricity for the plant. A fraction of the liquid stream from centrifugation is recycled back to the process, while the rest of the stream is sent to a wastewater treatment system.

Based on the equipment list generated from the process flow diagrams, the total capital cost for this plant in first-quarter 1990 dollars is \$141.24 MM. The annual capital charge rate is 20%, giving a capital charge of 48.3¢/gal. The variable operating cost (chemicals and feedstock) is 60.1¢/gal and the fixed operating cost (labor, taxes, and insurance) is 19.8¢/gal, giving a gross cost of production of 79.8¢/gal. When by-product credits (electricity) are included, the net cost of production is 73.4¢/gal. The total cost of production for the denatured fuel is 121.7¢/gal.

To assign priority to research issues, a sensitivity analysis was performed on major process variables and assumptions. The results for some of the major technical parameters that have a significant impact on ethanol cost are shown in Figure 1-1. This sensitivity analysis varied only one parameter while holding the other parameters at their base case values. The bars show the percent deviation of ethanol cost from the base case value of 121.7¢/gal, when the indicated changes are made from the base case values shown on the bottom of the figure. Particularly evident is the effect of nutrient requirements and SSF ethanol yield on the cost of ethanol. Figure 1-2 shows the impact of some issues that are not directly related to conversion technology, such as plant size and feedstock cost. Again, all these variables have a major impact on the cost of ethanol ranging from 5% to 15% or 6¢ to 18¢/gal of ethanol. When the effects of multiple process improvement are considered, the cost of ethanol drops to 66.5¢/gal.

Based on the results of the sensitivity analysis, future work should focus on strengthening understanding of the base case process, developing and improving the technology for biomass conversion, and continually analyzing and updating the process design. Understanding the base case process will involve efforts in four major areas: in-house integration research, vendor testing, subcontracted work to



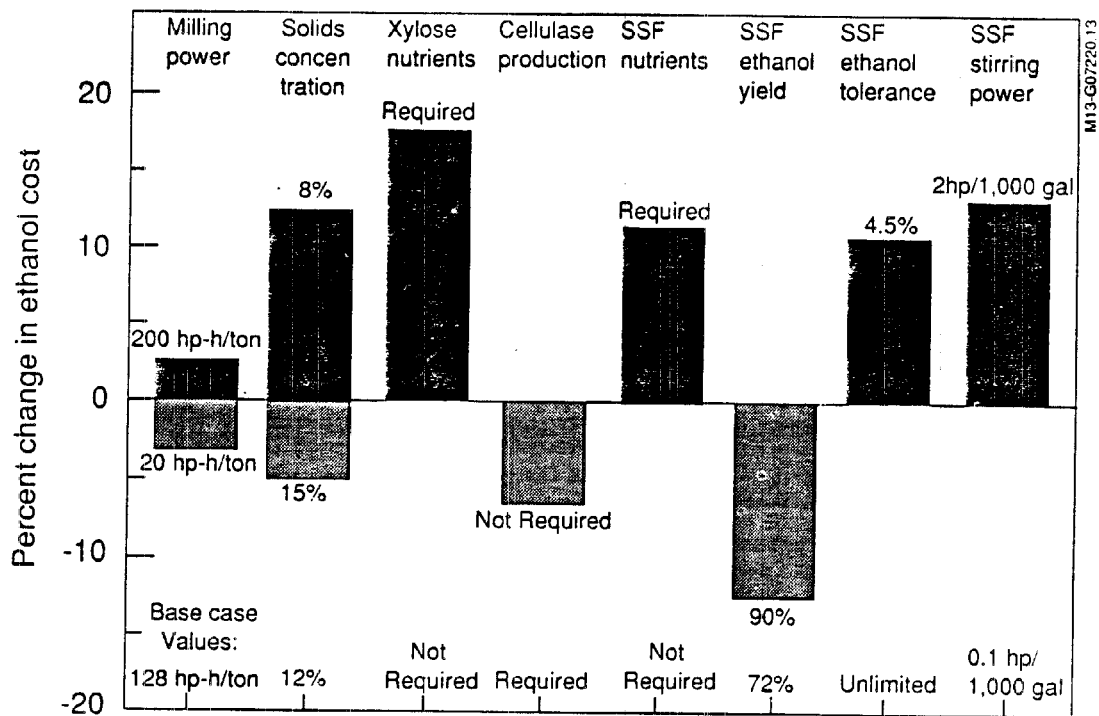


Figure 1-1. Percent change in ethanol cost caused by deviations of major technical parameters from their base case values

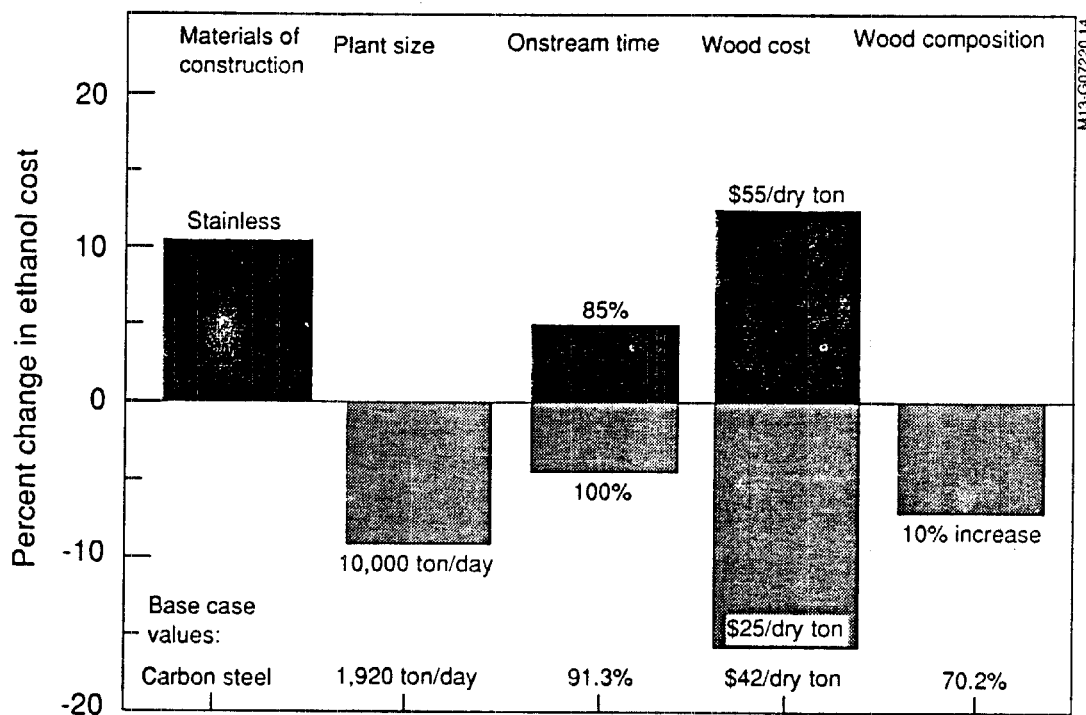


Figure 1-2. Percent change in ethanol cost caused by deviations of major process parameters from their base case values

engineering companies, and in-house pilot plant operations. The purpose of the integration research effort is to verify the performance of a fully chemically integrated system, which means testing the performance of pretreated feedstock and hydrolyzate from prehydrolysis through the back end of the plant as specified in the current process diagrams. Other issues should also be addressed, such as nutrient requirements, seedcultures, material balance closure, and process water recycle. Vendor testing is required to verify operation and performance data on equipment specified in the current process design. Equipment to be tested includes: mills, prehydrolysis and impregnation reactors, slurry pumps, large fermentation processes, distillation columns, lignin separation equipment, and boilers. Subcontracted work to engineering companies should examine such issues as materials of construction and waste treatment design. Pilot plant operation is required to verify the performance of the integrated process on a larger scale and to obtain information on process reliability and scale-up data for design of larger plants.

Although the primary goal of the above work is to verify the current technology, work must also progress on improving the technology. Specific areas to consider are improving ethanol yields, decreasing fermentation rates, increasing solids concentrations, eliminating or reducing seed fermentation and cellulase production requirements via recycle technology, and improving reactor designs and technologies. Because of the above tasks, periodic process analysis must also continue as new information is available in order to monitor research progress and identify areas for further research.

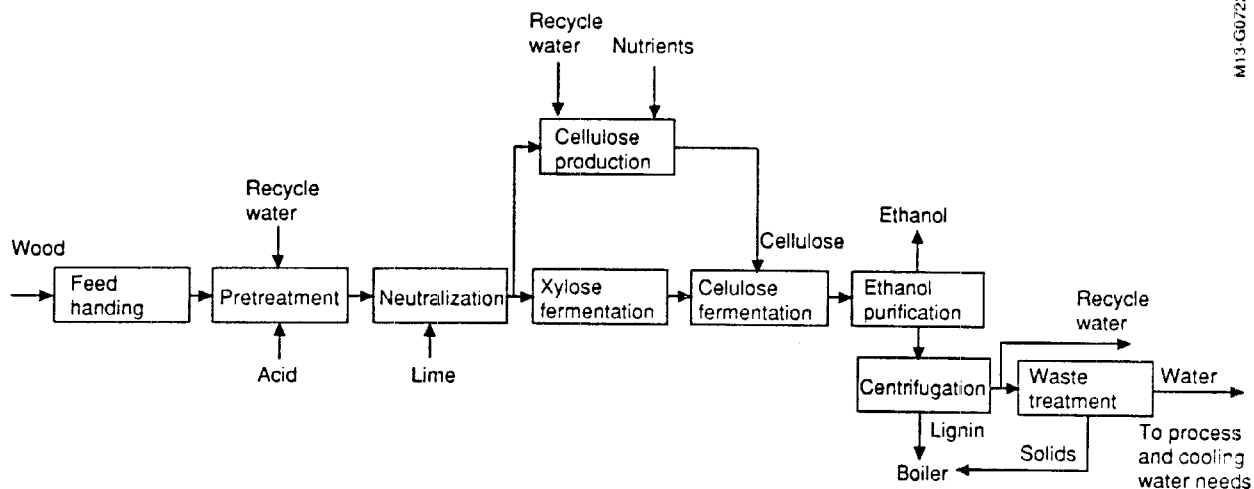
## Section 2.0

### Introduction

The purpose of this study was to perform a technical and economic analysis of an enzymatic biomass-to-ethanol process using SSF for conversion of cellulose to ethanol. Lignocellulosic biomass is an attractive feedstock for ethanol production because it is inexpensive and available in large quantities. For this analysis the feedstock was aspen wood, which is assumed to represent a wood energy crop that would be available in large quantities.

Wood is composed of three major components: cellulose, a polymer of glucose; xylan, a polymer of the five-carbon sugar xylose; and lignin, a phenolic compound. It is the carbohydrate components of the wood, cellulose, and xylan that are converted to ethanol. Cellulose is first enzymatically degraded to glucose by the enzyme cellulase, and then converted to ethanol by yeast. Xylan is degraded to xylose by a dilute sulfuric acid process and subsequently converted to ethanol by a microorganism. Lignin may have value as a chemical precursor, as an adhesive, or as a fuel additive. However, there is currently no viable technology for using the lignin. Thus, in this analysis it is carried through the process and burned to supply the plant with steam and electricity.

Based on new experimental data and current operating experience in the corn-to-ethanol industry, a new design was developed that is different from our previous work (Wright et al. 1988). The major steps for the enzymatic biomass-to-ethanol process are shown in Figure 2-1. From this concept, detailed



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Figure 2-1. Overall block flow diagram for the enzymatic biomass-to-ethanol process

flowsheets and material and energy balances were developed and used to generate equipment specifications and costs. An economic analysis was then performed to estimate the current cost of ethanol from biomass.

For the feed handling, pretreatment, neutralization, cellulase production, and xylose fermentation blocks, several technology options were considered. These options are discussed in Appendix A. Appendix A-1 discusses the different types of mills for biomass size reduction, from which a disk refiner was chosen for this study. A dilute sulfuric acid pretreatment was chosen from the various pretreatment options discussed in Appendix A-2. Appendix A-3 discusses the issues involved with separating the xylose from the lignin/cellulose particles remaining after pretreatment. Because of the difficulties involved in obtaining good separation of xylose from the remaining particles, this step was eliminated from the current design. Cellulase was produced using batch production techniques as opposed to fed-batch or continuous operation as discussed in Appendix A-4. The various options for xylose fermentation are simultaneous isomerization and fermentation (SFIX) (Lastick et al. 1989) with production of xylose isomerase (Lastick et al. 1986), use of xylose-fermenting yeast (Skoog and Hahn-Hagerdal 1988), use of xylose-fermenting *E. coli* (Ingram et al. 1987, Ingram and Conway 1988), or purchase of xylose isomerase for use with SFIX. These options are evaluated and discussed in Appendix A-5, from which *E. coli* was chosen for xylose fermentation. Appendix A-6 discusses nutrient requirements and cost for both xylose fermentation and SSF.

A simplified process concept flowsheet for the lignocellulosic biomass-to-ethanol process is shown in Drawing BESSF-022. Wood chips (aspen) are delivered to the plant and stored in a large pile. The wood is then conveyed to the disk refiner and reduced in size to approximately 2.0 to 3.0 mm. The milled chips are sent to a prehydrolysis reactor and treated with dilute sulfuric acid at 160°C for 10 min, which converts a large fraction of the wood xylans to xylose. The resulting mixture is flash cooled in a blowdown tank and the resulting vapor is sent to the distillation section of the plant to preheat the dilute ethanol stream. Recycled water is added to the mixture in the blowdown tank to reduce the solids concentration to a level that can be pumped to the neutralization tank. The sulfuric acid is then neutralized with lime and the resulting slurry, including the precipitated gypsum, is cooled and sent to xylose fermentation. The xylose is converted to ethanol by a genetically engineered *E. coli*. This dilute ethanol and cellulose/lignin stream is sent to cellulose fermentation where cellulose is converted to ethanol by the SSF process.

A small fraction of the neutralized stream is sent to the sterile feed tank and then pumped when needed to the cellulase production fermenters where the fungus *T. reesei* consumes the cellulose and produces cellulase. Nutrients and corn steep liquor are mixed and sterilized in a separate tank and then added to the cellulase fermenters. After the batch cellulase fermentation is complete, the broth is pumped to the cellulase hold tank and then pumped continuously to the SSF reactor. The cellulase enzyme catalyzes the hydrolysis of cellulose to glucose, which is then consumed by the yeast to produce ethanol all in the same reactor. The fermentation is carried out by a mixed culture of *Saccharomyces cerevisiae* and *Brettanomyces clauseni*.

The dilute ethanol stream (approximately 4.35 wt %) from the SSF reactors is sent to ethanol purification where a distillation and rectification process produces 95 wt % ethanol. Ethanol vapor in the fermentation off gases is condensed and added to the rectification column. The waste stream from the bottom of the beer column is sent to centrifugation to remove the lignin and unreacted solids, which are then sent to the boiler. A fraction of the liquid stream from the centrifuge is recycled back to the process and the rest is sent to waste treatment to be purified and then recycled back to the process water system. Utilities include a boiler, a turbogenerator, a boiler feed water system, off-site tanks, a process water system, plant and instrument air, fermentation air, a chilled water system, a low-pressure vent system, and a clean-in-place and chemical sterilization system.

In this study, heat and material balances were developed for the above lignocellulosic biomass-to-ethanol process based on the data and technical information presented in Section 3.0 and 4.0. Section 3.0 presents kinetic and rate data used to design the major processes in the plant. Section 4.0 gives the design basis for the plant, detailed process descriptions with design assumptions, the process utility summary, and the plant layout. The heat and material balances are used to specify equipment sizes and materials of construction. Costing information was obtained from literature references, previous work (Wright et al. 1988), studies by other groups (Badger Engineers, Inc. 1985, Stone and Webster, Inc. 1985), Chemcost, and vendor quotes. Section 5.0 presents the economic analysis for the base case plant and Section 6.0 discusses the optimum conditions for cellulase production and SSF performance. A sensitivity analysis of the important technical and economic parameters is presented in Section 7.0 and conclusions are given in Section 8.0. Recommendations for future research and development are presented in Section 9.0. References are given in Section 10.0. Appendix A is the discussion of various process options. Appendix B is a concise list of the technical data and assumptions used for process design and Appendix C is a list of design data (e.g., heat capacities and densities) Appendix D presents the spreadsheet model that was developed to analyze the process. Appendix E contains the process flowsheets and Appendix F is a complete equipment list with costs.

This process was designed according to our current knowledge and understanding of the biomass-to-ethanol process. Some assumptions have been made about the way the process operates; however, all yield data for xylose fermentations, SSF, and cellulase production were taken from laboratory experiments. The analysis was done to determine the areas of future research and development that will lead to the greatest reductions in the price of ethanol.

## Section 3.0

### Process Fundamentals

#### 3.1 Pretreatment with Dilute Sulfuric Acid

Pretreatment with dilute sulfuric acid is an important part of the overall biomass-to-ethanol process. This step hydrolyzes most of the xylan to xylose and a small amount of the cellulose to glucose. During pretreatment, some of the xylose is degraded to furfural and some of the glucose is degraded to hydroxymethylfurfural (HMF). The pretreatment process "opens up" the biomass, making the remaining cellulose more susceptible to enzymatic hydrolysis.

Previous work by Grohmann et al. (1985) has shown that pretreatment conditions of 0.85 wt % sulfuric acid, a temperature of 160°C, and a residence time of 10.0 min gives good xylan to xylose yields and makes the remaining cellulose digestible by cellulase. These results were obtained on low solid slurries (10 wt % solids) in a Parr (1.0-L batch) reactor. The substrate was aspen wood chips that were knife milled through a 1/8-in. rejection screen. Subsequent work by Grohmann et al. (1986) at the same pretreatment conditions has shown that solids concentration (up to 40 wt %) has no effect on enzyme digestibility.

Models for xylan hydrolysis are available from several authors (Grohmann et al. 1985, Kwarteng 1983, Kim and Lee 1987), but no one has been able to describe experimental results adequately. Therefore, in this study, experimental results obtained by Grohmann et al. (1986) at conditions described above were used for xylose (80% conversion of xylan to xylose) and furfural (13% conversion of xylan to furfural) yields. Although the data were obtained from a batch reactor, they are assumed to apply to the continuous plug flow reactor used in this analysis.

Cellulose hydrolysis and glucose degradation were determined from kinetic expressions and parameters available in the literature (Wright and d'Agincourt 1984). Cellulose consists of a crystalline fraction that is difficult to hydrolyze with dilute acid and an amorphous fraction that is more readily hydrolyzed (Wright and d'Agincourt 1984). For this study, the cellulose component of the feedstock was assumed to be all crystalline cellulose.

The hydrolysis of crystalline cellulose is described by the following reaction:



where  $k_1$  is the appropriate kinetic rate constant. At high temperatures (200°-260°C) 5-HMF further degrades to tars (Wright and d'Agincourt 1984), but in this study the relatively low temperature of pretreatment (160°C) is assumed to not produce tars.

The rate equations for crystalline cellulose and glucose are:

$$\frac{dC_c}{dt} = -k_1 C_c,$$

$$\frac{dG}{dt} = k_1 C_c - k_2 C_g,$$

where

$C_c$  = crystalline cellulose concentration expressed as a fraction of potential moles of glucose,  
 $C_g$  = glucose concentration expressed as a fraction of potential moles of glucose,  
 $k_1$  = crystalline cellulose degradation rate constant ( $\text{min}^{-1}$ ),  
 $k_2$  = glucose degradation rate constant ( $\text{min}^{-1}$ ), and  
 $t$  = time (min).

These equations are integrated to yield the following equations:

$$C_c = \exp(-k_1 t),$$

$$C_g = \frac{k_1}{k_2 - k_1} [\exp(-k_1 t) - \exp(-k_2 t)].$$

HMF is then determined by a mass balance as follows:

$$C_h = 1 - C_c - C_g.$$

where

$C_h$  = HMF concentration expressed as a fraction of potential moles of glucose.

The weight fraction conversion is obtained by multiplying the mole fraction by the following factors: 1.11 for glucose, 1.14 for xylose, 0.78 for HMF, and 0.73 for furfural.

The kinetic rate constants are determined from the following Arrhenius expression:

$$k_i = K_i(A)^{n_i} \exp\left(-\frac{E_i}{RT}\right).$$

Values for the parameters in the equation are defined as follows:

$A$  = acid concentration as calculated in the liquid phase (wt %),  
 $R$  = gas constant (cal/gmole-K),  
 $T$  = temperature (K),  
 $E_i$  = activation energy (cal/gmole),  
 $K_i$  = pre-exponential factor, and  
 $n_i$  = an exponent.

Values for the parameters are given in Table 3-1.

**Table 3-1. Kinetic Parameters for Dilute Acid Hydrolysis of Crystalline Cellulose**

i	$K_i$ $\text{min}^{-1}$	$n_i$	$E_i$ cal/mole	Source
1	$1.730 \times 10^{19}$	1.34	42900	Saeman 1945
2	$1.790 \times 10^{14}$	0.84	32506	McKibbins 1962

In this analysis, lignin is assumed to be unchanged by the prehydrolysis conditions.

Heat and mass transfer in both the impregnation and prehydrolysis reactors used in this study are assumed to be as good as the batch experimental apparatus used by Grohmann et al. (1986). These assumptions need to be tested for the actual reactors assumed in this study, particularly the impregnation of high-solids wood slurries by sulfuric acid.

### 3.2 Neutralization

After pretreatment, the slurry is flashed to atmospheric pressure, mixed with water, and sent to a vessel where the material is neutralized with lime with the resulting production of gypsum. Gypsum production and lime use are determined from the following equation:



Thus, 0.76 lb  $\text{Ca}(\text{OH})_2$  is used per pound of  $\text{H}_2\text{SO}_4$  and 1.76 lb gypsum is produced per pound of  $\text{H}_2\text{SO}_4$ . A fraction of the total gypsum produced remains in solution up to the saturation value at a given temperature, and the rest is insoluble gypsum. The possibility exists that as the temperature of the process stream increases, gypsum may plate out on some of the equipment. This might occur in the beer column reboiler, which is at a higher temperature than the temperature of the slurry when originally neutralized. In this analysis, it is assumed that plating out is not a problem. This assumption needs to be investigated further.

### 3.3 Xylose Fermentation via Genetically Engineered *E. coli*

The xylose-fermenting microorganism used in the study is a genetically engineered *E. coli* developed by L. Ingram and coworkers at the University of Florida. The addition of genes (alcohol dehydrogenase II and pyruvate decarboxylase from *Zyomonas mobilis*) *E. coli* TC4 results in ethanol as the main fermentation product.

This microorganism was donated to SERI for testing and evaluation. It was tested by D. Spindler (1989) in anaerobic batch cultures (5.0-L fermenters) for up to 2 days on a pure 8.0 wt % xylose solution with complete luria broth (LB) media (10.0 g/L tryptone, 5.0 g/L yeast extract, and 10.0 g/L NaCl). The pH of the fermentations was controlled at 5.5, 6.0, 6.5, and 7.0 with  $\text{NH}_3\text{OH}$  and the temperature was maintained at 37°C. The inoculum was 10% (v/v) and grown in an agitated aerobic shake flask on 2.0 wt % xylose in LB media for 12 h at 37°C. The pH was not controlled or measured during inoculum growth.

Data collected on ethanol yield and  $\text{NH}_3\text{OH}$  use are shown in Table 3-2. These data were fitted to a multiple regression curve, but the curve is not used in this analysis because of large residual errors.



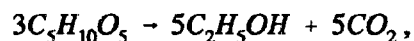
Instead, the performance at a pH of 6.5 and a residence time of 1.87 days is used. With these conditions, the xylose to ethanol yield is assumed to be 95% and  $\text{NH}_3\text{OH}$  consumption is 3.5 lb  $\text{NH}_3\text{OH}$ /gal of ethanol produced. For this study, these data are assumed to apply to any xylose concentration near or less than 8 wt % and to be applicable to a series of continuous stirred-tank fermenters. This latter assumption is appropriate, because a series of continuous stirred-tank fermenters resembles a plug flow reactor, which in turn can be described with batch data. For this analysis, it is also assumed that the use of recycle water would provide all the necessary nutrients to support microorganism growth. The temperature is 37°C and the pH was 6.5, the same as in the experimental results described above.

**Table 3-2. *E. coli* Xylose Fermentation Data for 8.0% Xylose**

Time (d)	pH	Ethanol Yield (%)	$\text{NH}_4\text{OH}$ Usage (lb/gal ethanol)
0.04	5.5	2.7	94.1
0.33	"	8.8	36.7
1.00	"	34.3	13.4
1.33	"	40.9	13.0
2.00	"	55.5	9.6
0.04	6.0	2.4	0.0
0.41	"	19.9	2.1
0.91	"	63.8	1.7
1.25	"	83.7	1.7
1.87	"	88.9	1.6
0.04	6.5	2.4	0.0
0.41	"	22.9	4.9
0.91	"	79.4	4.0
1.25	"	94.9	3.5
1.87	"	96.2	3.5
0.04	7.0	2.5	27.7
0.33	"	15.4	9.0
1.00	"	79.6	3.5
1.33	"	84.2	4.1
2.00	"	89.8	3.9

Source: Spindler 1989

The theoretical yield of ethanol and carbon dioxide from xylose is calculated from the following equation:



yielding 0.51 lb ethanol/lb xylose and 0.49 lb CO<sub>2</sub>/lb xylose. Also, any glucose in the feed is assumed converted to ethanol at the same yield as xylose and with the same theoretical conversion efficiencies as xylose. It is assumed that no microorganism growth occurred in the main fermenters.

The agitation energy is assumed to be 0.1 hp/1,000 gal. This assumption is based on technical information derived from visits to existing corn-to-ethanol plants and subcontracted work by Elander (1988) at Colorado State University (CSU). The heat of fermentation is estimated from the heat of reaction for xylose to ethanol (612 Btu/lb xylose), which is calculated from heat of combustion data (see Appendix C). This plus the agitation energy input is used to estimate the heat duty of the fermenters (1.82 MMBtu/h/fermenter, 14.6 MMBtu/h total heat duty). All other heat inputs or losses (i.e., evaporation and conduction) are ignored.

In this study, as per the experimental data and conditions discussed above, each seed fermenter supplied a 10% inoculum to each succeeding fermenter. The fermentation time is 12 h with a total cycle time of 1 day, and a starting xylose/glucose concentration of 2.0 wt %. Because the pH is not controlled, no base is used. Temperature is controlled at 37°C, and it is assumed that the use of recycled water would provide all nutrients required for microorganism growth. All xylose is assumed converted to cell mass at a yield of 0.5 lb cells/lb xylose. The stirring power and aeration rate is assumed to be 0.1 hp/1,000 gal and 0.2 vvm, respectively. The heat duty is estimated by assuming all the combustion energy of the substrates (glucose and xylose) is converted to heat (4.8 MMBtu/h).

### 3.4 Cellulase Production

Although cellulase enzymes are produced commercially by Genencor International, Inc., and by Novo Industries, their production methods and data are confidential. Cellulase has been produced in a 3000-L fermenter (Warzywoda et al. 1983), but the substrate was lactose. Also, Bevernitz and coworkers (1982) claimed to have achieved pilot plant results that compared favorably with laboratory scale fermenters, but no data were published and fermenter sizes were not specified. Moreover, the organisms used in this latter work were not the latest improved cellulase-producing strains.

Laboratory scale results for cellulase production from cellulose (primarily Solka Floc) using aerobic batch fermenters are compiled in Table 3-3. The rationale for using batch data, as opposed to fed-batch data, is discussed in Appendix A-4. The data shown in Table 3-3 were obtained for the most part at a temperature of 28°C and a pH of 4.8. The average yield and residence time for the data are shown at the bottom of Table 3-3.

In our study, the feed contained xylose (4.0%) as well as cellulose (7.6%). However, Mohagheghi et al. (1988) showed that substitution of cellulose by xylose (less than 50%) with maintenance of the same overall carbohydrate concentration has no effect on cellulase production performance. Therefore, the average values for performance parameters shown at the bottom of Table 3-3 were used for the batch fermenters in this study, except that in this study, performance parameters, such as yield, were based on total carbohydrate (cellulose and xylose) rather than just cellulose. Because information is not available on *T. reesei* growth stoichiometry, carbon dioxide production is ignored and the cells are assumed to grow to 20 g/L in the main fermenters.

**Table 3-3. Cellulase Production Data for *Trichoderma reesei*<sup>a</sup>**

Cellulose Concentration	Specific Activity	Residence Time	Cellulase Yield	Source
(g/L)	(IU/g enzyme)	(d)	(IU/g cellulose)	
50	1220	6.00	240	Sheir-Ness and Montecourt 1984
50	660	8.00	288	Tangnu et al. 1981
50		4.00	130	Hendy et al. 1982
70	800	5.00	286	Durand et al. 1988
50		7.00	240	Watson and Nelligan 1983
50		7.00	160	Schell et al. 1990
40	100	2.75	69	Mohagheghi et al. 1988

Average residence time: 5.7 d

Average cellulase yield: 202 IU/g

<sup>a</sup> Strain Rut C-30 except for Durand et al. (1988), who used strain CL-847; and Schell et al. (1990), who used strain L-27.

In this study, the temperature is 28°C and the pH is 4.8. The NH<sub>4</sub>OH consumption is 0.045 lb/lb of carbohydrate in the feed, determined from the average of two experimental values in the literature (Sternberg and Durval 1979, Watson and Nelligan 1983). The media composition used in our study is taken from the work of Wilke and Blanch (1985) and is shown in Table 3-4.

**Table 3-4. Media for Cellulase Production (Wilke and Blanch 1985)**

Component	Amount (g/L)
(NH <sub>4</sub> ) <sub>2</sub> SO <sub>4</sub>	1.4
KH <sub>2</sub> PO <sub>4</sub>	2.0
MgSO <sub>4</sub> ·7H <sub>2</sub> O	0.3
CaCl <sub>2</sub> ·2H <sub>2</sub> O	0.4
Tween 80	0.2
Corn steep liquor	15.0

In this analysis, the cellulase fermenter agitation power and air requirements are determined from correlations and literature data on *T. reesei* as follows. The specific growth rate for *T. reesei* was determined to be .0415/h from the data of Mohagheghi et al. (1988). From the data of Ryu et al. (1979), the specific oxygen uptake rate (mM O<sub>2</sub>/g-h) was determined from the specific growth rate. From the specific oxygen uptake rate and the assumed maximum cell concentration of 20 g/L, the maximum oxygen uptake rate (N<sub>a</sub>, mM O<sub>2</sub>/L-h) was estimated as 42 mM O<sub>2</sub>/L-h.

The oxygen uptake rate (N<sub>a</sub>) is a function of the oxygen transfer coefficient (k<sub>L</sub>a) and the equilibrium and dissolved bulk oxygen concentrations in the fermenter. This relationship for large fermenters is (Wang et al. 1979):

$$N_a = \frac{k_L a [(C^*_{\text{b}} - C_L) - (C^*_{\text{t}} - C_L)]}{\ln[(C^*_{\text{b}} - C_L)/(C^*_{\text{t}} - C_L)]}$$

where

$$\begin{aligned} C^*_{\text{b}} &= \text{equilibrium oxygen concentration at the bottom of the fermenter,} \\ C^*_{\text{t}} &= \text{equilibrium oxygen concentration at the top of the fermenter,} \\ C_L &= \text{dissolved oxygen concentration.} \end{aligned}$$

The dissolved bulk oxygen concentration is assumed to be 20% of the average of  $C^*_{\text{b}}$  and  $C^*_{\text{t}}$ , where  $C^*_{\text{b}}$  and  $C^*_{\text{t}}$  are a function of the volumetric air rate and fermenter pressure. The coefficient,  $k_L a$ , is a function of the stirring power and volumetric air rate. Thus, given a fermenter pressure, the maximum oxygen uptake establishes a set of possible stirring powers and volumetric air rates. The desired stirring power and volumetric air rate are those that result in minimum overall electricity use.

Calculation of the desired stirring power and volumetric air rate from the maximum oxygen uptake rate and assumed fermenter pressure proceeded as follows. For the maximum oxygen uptake rate, an initial guess was made of the volumetric air rate to the fermenter. From this volumetric air rate, the oxygen uptake rate, the concentration of oxygen in the inlet air, and the fermenter pressure, it is possible to determine the oxygen concentration in the off gas. This concentration, along with the inlet oxygen concentration, is used to calculate the equilibrium oxygen concentration at the top ( $C^*_{\text{t}}$ ) and bottom ( $C^*_{\text{b}}$ ) of the fermenter using Henry's Law. Knowing  $C^*_{\text{b}}$  and  $C^*_{\text{t}}$  along with  $N_a$ , and assuming  $C_L$  to be 20% of the average of  $C^*_{\text{b}}$  and  $C^*_{\text{t}}$ , it is possible to estimate  $k_L a$  from the equation given above.

A correlation that relates the oxygen transfer coefficient ( $k_L a$ ) to stirring power and the volumetric air rate is given by Wang et al. (1979) for a non-Newtonian fermentation broth as:

$$k_L a = 8.3 \left( \frac{P_a}{V} \right)^{0.33} V_s^{0.56},$$

where

$$\begin{aligned} P_a/V &= \text{gassed stirring power input per unit volume (hp/1,000L), and} \\ V_s &= \text{superficial gas velocity (cm/min).} \end{aligned}$$

The value for  $V_s$  is calculated from the volumetric air rate assumed for the initial calculation of  $k_L a$  and the fermenter dimensions. Knowing  $k_L a$  and  $V_s$ , it is possible to calculate a value for  $P_a/V$ . From the volumetric air rate and  $P_a/V$ , the total compression and stirring power can be calculated.

The power required for air compression ( $P_c$ ) is given by (Perry and Chilton 1973):

$$P_c = 0.0044 p_1 Q_1 \ln \left( \frac{p_2}{p_1} \right),$$

where

$$\begin{aligned} p_1 &= \text{inlet gas pressure (atm),} \\ p_2 &= \text{outlet gas pressure (atm), and} \\ Q_1 &= \text{inlet gas flow rate (ft}^3\text{/min).} \end{aligned}$$

At this point, a total energy requirement is established for a set of values for stirring power and volumetric air flow that satisfy the maximum oxygen uptake rate. The entire procedure is then repeated until a set of values for stirring power and air flow rate are found that satisfy the maximum oxygen uptake required while requiring a minimum total energy input.

Using the above procedure with an assumed fermenter pressure of 45 psig, the optimum volumetric air flow rate for cellulase production was calculated to be 0.15 vvm and the mixing power for a gassed fermenter is calculated to be 1.18 hp/1,000 gal.

The heat duty for each cellulase production fermenter results from metabolic activities of the cells, plus agitation energy input, plus gassed power input minus evaporation losses. Conduction losses are assumed to be zero. Metabolic heating ( $Q_f$ , kcal/L-h) is determined from the following formula (Wang et al. 1979):

$$Q_f = 0.12 N_A.$$

All agitation energy is assumed to be absorbed by the fluid as heat. Gassed compression power input per unit volume ( $P_g/V$ ) is estimated from the following formula (Atkinson and Mavituna 1983):

$$\frac{P_g}{V} = \left( \frac{QD}{V} \right) \left( \frac{RT}{M} \right) \ln \left( \frac{p_o}{p} \right),$$

where

Q is gas flow rate (scf/min),  
d is gas density (lb/cf),  
R is the gas constant,  
T is gas temperature (K),  
M is gas molecular weight (g/gmole),  
 $p_o$  is pressure at the sparger (atm), and  
p is absolute pressure in the fermenter (atm).

Evaporation losses are determined from the change in enthalphy of the entering and exiting gas streams. From the calculations, the heat duty is 13.1 MMBtu/h.

Inoculum volumes used in the studies shown in Table 3-3 ranged from 2% to 10% (v/v). An average value of 5% is used in this study. In addition, based on studies by Wilke and Blanch (1985) that used a 1% cellulose concentration, in this study a 1.0% cellulose plus xylose concentration with the media composition shown in Table 3-4 is assumed for the final and largest cellulase production seed fermenter. For all other seed fermenters, 1% glucose with media composition shown in Table 3-4 is assumed and the cell yield for all seed fermenters is assumed to be 0.5 g/g substrate. The cycle time for all batch seed fermenters is 4 days based on the inoculum procedure of Wilke and Blanch (1985). The temperature is 28°C and the pH is assumed controlled to 4.8. Base usage is assumed to be small and ignored because there are no data on base consumption during seed culture growth. The air flow rate for all seed fermenters is assumed to be 0.2 vvm and the agitation requirements are assumed to be 0.5 hp/1,000 gal for the final seed fermenter and 1.0 hp/1,000 gal for all other seed fermenters. The heat duty for the seed fermenters is estimated as 5% of the main fermenters.

### 3.5 Simultaneous Saccharification and Fermentation (SSF)

SSF is a promising method of producing ethanol from lignocellulosics. The process uses cellulase enzyme to saccharify cellulose to glucose, which is converted to ethanol by yeast, all in the same reactor. The immediate removal of the glucose prevents inhibition of the cellulase enzyme. SSF has been studied by a number of researchers (Takagi et al. 1977, Meyers 1978, Ghosh et al. 1982, Deshpande et al. 1983, Spindler et al. 1988, Huang and Chen 1988, Szczodrak 1988, Szczodrak 1989, Spindler et al. 1989a, Spindler et al. 1989b) and ethanol yields vary depending on enzyme loading and quality, initial cellulose concentration, cellulose digestibility, and fermenting organisms.

Experimental ethanol yields obtained by Spindler (1989) for SSF with uncontrolled pH (pH drops from an initial value of 5.0 to approximately 3.5), are given in Table 3-5. These numbers were obtained with 100-mL anaerobic shake flask experiments at 37°C, with either pretreated (with dilute sulfuric acid) wheat straw or *Populus* as the substrate at a 7.5% cellulose concentration and with a complex media (yeast extract and peptone). The pretreated substrate was washed to remove acid and other by-products before it was used in the SSF experiments. The cellulase used was a commercial preparation obtained from Genencor International, defined by Spindler et al. (1989a) as Genencor batch II, as opposed to an earlier preparation obtained from Genencor, which has been labeled Genencor batch I. Genencor batch I had superior SSF performance when compared to batch II (Spindler et al. 1988), even though the manufacturer filter paper assay (150 IU/mL) was the same (Genencor 150 L Product Sheet). This indicates that more work is needed to define quality standards for cellulase when used in SSF. Also tested in the studies by Spindler (1989) were two different types of yeast: (1) *S. cerevisiae* D<sub>5</sub>A and (2) a mixed culture of *S. cerevisiae* D<sub>5</sub>A and *B. clausenii* Y1414.

In order to calculate ethanol yield at different conditions, the data shown in Table 3-5 is linearly regressed as a function of time and enzyme loading to the following equation:

$$Y_2 = a + bt + ct^2 + dL + eL^2 + fLt,$$

where

- $Y_2$  = the theoretical ethanol yield at a 7.5% cellulose concentration with Genencor batch II cellulase (%),
- $L$  = cellulase loading (IU/g cellulose),
- $t$  = time (d), and
- $a, b, c, d, e, f$  = coefficients determined from the least squares multiple regression.

The coefficients are given in Table 3-6. These are good for enzyme loadings from 7-26 IU/g cellulose and between 2-5 days for *S. cerevisiae* with wheat straw or 2-8 days for the other three culture-substrate combinations. Plots of the predicted ethanol yields as a function of time, yeast culture type, and substrate type for a cellulase loading of 7 IU/g cellulose is shown in Figure 3-1. Note that a mixed culture has higher ethanol yields than *S. cerevisiae* alone and that wheat straw is better than *Populus*.

Table 3-5. SSF Performance Data for Different Substrates as a Function of Enzyme Loading and Time

Substrate	Cellulose Concentration (g/L)	Cellulose Loading (IU/g)	Cellulase Type	Ethanol		Residence Time (d)	Organism
				Yield (%)	Concentration (g/L)		
Wheat Straw	75	7	Genencor II	16.0	6.8	1	<i>S. cerevisiae</i>
"	"	"	"	28.2	12.0	2	"
"	"	"	"	35.8	15.2	3	"
"	"	"	"	42.4	18.0	4	"
"	"	"	"	44.8	19.0	5	"
"	"	13	"	25.4	10.8	1	"
"	"	"	"	39.6	16.8	2	"
"	"	"	"	47.1	20.0	3	"
"	"	"	"	54.6	23.2	4	"
"	"	"	"	55.3	23.5	5	"
"	"	26	"	40.0	17.0	1	"
"	"	"	"	61.2	26.0	2	"
"	"	"	"	69.5	29.5	3	"
"	"	"	"	76.3	32.4	4	"
"	"	"	"	76.3	32.5	5	"
"	"	7	"	28.2	12.0	0.75	Mixed   <sup>a</sup>
"	"	"	"	46.2	20.0	1.8	"
"	"	"	"	53.0	23.0	3	"
"	"	"	"	56.5	24.0	4	"
"	"	"	"	61.7	26.0	5	"
"	"	"	"	64.1	27.0	6	"
"	"	"	"	65.9	28.0	7	"
"	"	"	"	65.9	28.0	8	"
"	"	13	"	36.7	16.0	0.75	"
"	"	"	"	55.6	24.0	1.8	"

Table 3-5. SSF Performance Data for Different Substrates as a Function of Enzyme Loading and Time (Continued)

Substrate	Cellulose Concentration (g/L)	Cellulose Loading (IU/g)	Cellulase Type	Ethanol		Residence Time (d)	Organism
				Yield <sub>i</sub> (%)	Concentration (g/L)		
Wheat Straw	75	13	Genencor II	60.1	26.0	3	Mixed I
"	"	"	"	63.6	27.0	4	"
"	"	"	"	66.9	28.0	5	"
"	"	"	"	68.8	29.0	6	"
"	"	"	"	71.6	30.0	7	"
"	"	"	"	72.5	31.0	8	"
"	"	26	"	37.7	16.0	0.75	"
"	"	"	"	54.2	23.0	1.8	"
"	"	"	"	61.2	26.0	3	"
"	"	"	"	65.9	28.0	4	"
"	"	"	"	68.3	29.0	5	"
"	"	"	"	70.7	30.0	6	"
"	"	"	"	71.8	31.0	7	"
"	"	"	"	75.4	32.0	8	"
Populus	"	7	"	7.1	3.0	1	<i>S. cerevisiae</i>
"	"	"	"	12.2	5.0	2	"
"	"	"	"	19.8	8.0	3	"
"	"	"	"	23.6	10.0	4	"
"	"	"	"	29.7	13.0	5	"
"	"	"	"	34.9	15.0	6	"
"	"	"	"	39.6	17.0	7	"
"	"	"	"	42.4	18.0	8	"
"	"	13	"	10.8	5.0	1	"
"	"	"	"	19.8	8.0	2	"
"	"	"	"	30.1	13.0	3	"



Table 3-5. SSF Performance Data for Different Substrates as a Function of Enzyme Loading and Time (Continued)

Substrate	Cellulose Concentration (g/L)	Cellulose Loading (IU/g)	Cellulase Type	Ethanol		Residence Time (d)	Organism
				Yield (%)	Concentration (g/L)		
Populus	75	13	Genencor II	37.7	16.0	4	<i>S. cerevisiae</i>
"	"	"	"	45.9	20.0	5	"
"	"	"	"	52.8	22.0	6	"
"	"	"	"	58.4	25.0	7	"
"	"	"	"	61.9	26.0	8	"
"	"	19	"	14.1	6.0	1	"
"	"	"	"	27.3	12.0	2	"
"	"	"	"	40.5	17.0	3	"
"	"	"	"	51.8	22.0	4	"
"	"	"	"	58.4	25.0	5	"
"	"	"	"	66.9	28.0	6	"
"	"	"	"	72.5	31.0	7	"
"	"	"	"	74.4	32.0	8	"
"	"	26	"	8.8	8.0	1	"
"	"	"	"	33.9	14.0	2	"
"	"	"	"	49.9	21.0	3	"
"	"	"	"	61.2	26.0	4	"
"	"	"	"	70.7	30.0	5	"
"	"	"	"	75.4	32.0	6	"
"	"	"	"	78.2	33.0	7	"
"	"	"	"	80.1	34.0	8	"
"	"	7	"	12.2	3.0	1	Mixed I <sup>a</sup>
"	"	"	"	23.6	5.0	2	"
"	"	"	"	31.8	8.0	3	"
"	"	"	"	38.9	10.0	4	"

Table 3-5. SSF Performance Data for Different Substrates as a Function of Enzyme Loading and Time (Continued)

Substrate	Cellulose Concentration (g/L)	Cellulose Loading (IU/g)	Cellulase Type	Ethanol		Residence Time (d)	Organism
				Yield (%)	Concentration (g/L)		
Populus	75	7	Genencor II	46.2	13.0	5	Mixed I
"	"	"	"	49.6	15.0	6	"
"	"	"	"	51.8	17.0	7	"
"	"	"	"	53.8	18.0	8	"
"	"	13	"	16.8	5.0	1	"
"	"	"	"	31.1	8.0	2	"
"	"	"	"	42.4	13.0	3	"
"	"	"	"	51.8	16.0	4	"
"	"	"	"	61.2	20.0	5	"
"	"	"	"	66.9	22.0	6	"
"	"	"	"	67.8	25.0	7	"
"	"	"	"	68.7	26.0	8	"
"	"	19	"	24.7	6.0	1	"
"	"	"	"	34.2	12.0	2	"
"	"	"	"	45.9	17.0	3	"
"	"	"	"	56.5	22.0	4	"
"	"	"	"	65.9	25.0	5	"
"	"	"	"	70.7	28.0	6	"
"	"	"	"	73.7	31.0	7	"
"	"	"	"	74.2	32.0	8	"
"	"	26	"	25.9	8.0	1	"
"	"	"	"	35.3	14.0	2	"
"	"	"	"	48.3	21.0	3	"
"	"	"	"	59.4	26.0	4	"
"	"	"	"	70.7	30.0	5	"

Table 3-5. SSF Performance Data for Different Substrates as a Function of Enzyme Loading and Time (Concluded)

Substrate	Cellulose Concentration (g/L)	Cellulase Loading (IU/g)	Cellulase Type	Ethanol		Residence Time (d)	Organism
				Yield (%)	Concentration (g/L)		
Populus	75	26	Genencor II	73.7	32.0	6	Mixed <sup>a</sup>
"	"	"	"	74.9	33.0	7	"
"	"	"	"	75.4	34.0	8	"

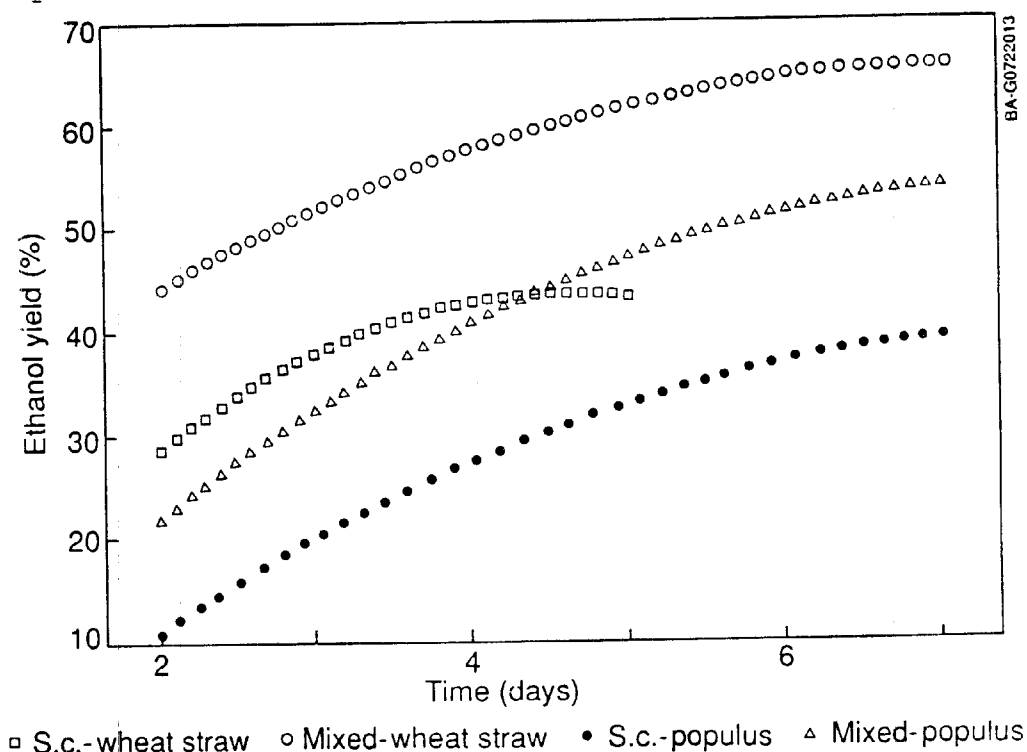
Source: Spindler 1989  
<sup>a</sup> *S. cerevisiae* and *B. clausenii*

**Table 3-6. Regression Coefficients for Yield at 7.5% Cellulose**

Culture	Substrate	a	b	c	d	e	f
<i>S. cerevisiae</i>	Wheat straw	-15.35	19.56	-2.18	1.85	-0.014	0.086
Mixed I <sup>a</sup>	Wheat straw	11.08	11.71	-0.82	2.25	-0.055	-0.006
<i>S. cerevisiae</i>	<i>Populus</i>	-26.10	11.62	-0.84	2.36	-0.053	0.229
Mixed I	<i>Populus</i>	-22.17	15.06	-1.02	2.89	-0.068	0.077

<sup>a</sup>*S. cerevisiae* and *B. clausenii*

In this analysis, a mixed culture is used with a hardwood feedstock (*Populus*). In addition, the cellulase is assumed to be Genencor batch I. Because the regressed function for  $Y_2$  is obtained with data using Genencor batch II, a modification to  $Y_2$  is required to determine the ethanol yield  $Y_1$  with Genencor batch I cellulase. The modification is done using the data of Table 3-7 (Spindler et al. 1988), which shows ethanol yield with Sigmacell 50 at a 7.5% cellulose concentration as a function of the two cellulase types, cellulase loadings from 7-26 IU/g cellulose, and the two cultures previously defined. Temperature, pH conditions, and media are the same as for the data in Table 3-5. These data are linearly regressed using the form of the equation given above and yielding the coefficients shown in Table 3-8.



**Figure 3-1. Ethanol yield from SSF at 7 IU/g cellulose for various substrates using *S. cerevisiae* (S.c.) and a mixed culture**

Table 3-7. SSF Performance Data for Different Genencor Cellulase Batches

Substrate	Cellulose Concentration (g/L)	Cellulase Loading (U/g)	Cellulase Type	Ethanol Yield (%)	Residence Time (d)	Organism
Sigmatell 50	75	7	Genencor I	35.1	2	<i>S. cerevisiae</i>
"	"	13	"	45.0	"	"
"	"	26	"	56.6	"	"
"	"	7	"	49.5	4	"
"	"	13	"	61.1	"	"
"	"	26	"	70.1	"	"
"	"	7	"	69.2	7	"
"	"	13	"	78.2	"	"
"	"	26	"	81.8	"	"
"	"	7	"	46.7	2	Mixed   <sup>a</sup>
"	"	13	"	55.7	"	"
"	"	26	"	60.2	"	"
"	"	7	"	70.1	4	"
"	"	13	"	75.5	"	"
"	"	26	"	75.5	"	"
"	"	7	"	85.4	7	"
"	"	13	"	88.1	"	"
"	"	26	"	88.1	"	"
"	"	7	Genencor II	16.0	1	<i>S. cerevisiae</i>
"	"	"	"	24.5	2	"
"	"	"	"	32.5	3	"
"	"	"	"	40.5	4	"
"	"	"	"	47.1	5	"
"	"	"	"	49.9	6	"
"	"	"	"	50.6	7	"

Table 3-7. SSF Performance Data for Different Genencor Cellulase Batches (Continued)

Substrate	Cellulose Concentration (g/L)	Cellulase Loading (IU/g)	Cellulase Type	Ethanol Yield (%)	Residence Time (d)	Organism
Sigmatell 50	75	13	Genencor II	22.3	1	<i>S. cerevisiae</i>
"	"	"	"	32.0	2	"
"	"	"	"	41.5	3	"
"	"	"	"	50.6	4	"
"	"	"	"	55.6	5	"
"	"	"	"	59.6	6	"
"	"	"	"	60.1	7	"
"	"	26	"	28.3	1	"
"	"	"	"	41.0	2	"
"	"	"	"	48.3	3	"
"	"	"	"	56.5	4	"
"	"	"	"	58.6	5	"
"	"	"	"	64.1	6	"
"	"	"	"	64.8	7	"
"	"	7	"	21.7	1	Mixed <sup>a</sup>
"	"	"	"	34.9	2	"
"	"	"	"	44.6	3	"
"	"	"	"	51.8	4	"
"	"	"	"	58.4	5	"
"	"	"	"	62.2	6	"
"	"	"	"	63.6	7	"
"	"	13	"	27.3	1	"
"	"	"	"	41.7	2	"
"	"	"	"	49.9	3	"
"	"	"	"	58.4	4	"

Table 3-7. SSF Performance Data for Different Genencor Cellulase Batches (Concluded)

Substrate	Cellulose Concentration (g/L)	Cellulase Loading (IU/g)	Cellulase Type	Ethanol Yield (%)	Residence Time (d)	Organism
Sigmacell 50	75	13	Genencor II	64.5	5	Mixed I <sup>a</sup>
"	"	"	"	65.9	6	"
"	"	"	"	65.9	7	"
"	"	26	"	33.3	1	"
"	"	"	"	49.0	2	"
"	"	"	"	54.2	3	"
"	"	"	"	63.1	4	"
"	"	"	"	66.9	5	"
"	"	"	"	67.8	6	"
"	"	"	"	68.3	7	"

Source: Spindler et al. 1988, Spindler 1989  
<sup>a</sup> *S. cerevisiae* and *B. clausenii*

**Table 3-8. Regression Coefficients for Yield for Different Genencor Cellulases**

Culture	Cellulase	a	b	c	d	e	f
<i>S. cerevisiae</i>	Genencor I	-5.44	11.22	-0.39	3.27	-0.056	0.102
Mixed I <sup>a</sup>	Genencor I	4.62	17.60	-1.05	2.27	-0.044	-0.101
<i>S. cerevisiae</i>	Genencor II	-13.47	13.31	-0.86	2.56	-0.053	-0.012
Mixed I	Genencor II	-5.04	16.58	-1.14	1.84	-0.031	-0.074

<sup>a</sup> *S. cerevisiae* and *B. clausenii*

The yield with Genencor I ( $Y_I$ ) is obtained by ratioing Sigmacell 50 yields for the two enzymes at the same enzyme loading and multiplying this ratio by the yield for Genencor II ( $Y_{II}$ ) according to the following formula:

$$Y_I = Y_{II} \left( \frac{Y_I}{Y_{II}} \right),$$

where

$Y_I$  = the ethanol yield with Genencor I and 7.5% Sigmacell 50 (%),  
 $Y_{II}$  = the ethanol yield with Genencor II and 7.5% Sigmacell 50 (%).

The above discussion assumed a cellulose concentration of 7.5%. If a different cellulose concentration is required, the yield previously determined must be further modified. The data in Table 3-9 (Spindler et al. 1988) present ethanol yield for Sigmacell 50 at different substrate concentrations. Conditions of temperature, pH, and media are the same as that for the data in Table 3-5. These data are linearly regressed using the equation shown below:

$$Y_{cc} = a + bC + ct + dCt,$$

where

C = cellulose concentration (%),  
t = time (d), and  
a,b,c,d = regression coefficients.

The coefficients are given in Table 3-10 and are valid for a cellulose concentration between 7.5%-15.0%.

**Table 3-10. Regression Coefficients for Yield versus Cellulose Concentration**

Culture	a	b	c	d
<i>S. cerevisiae</i>	43.91	-1.20	6.20	-0.36
Mixed I <sup>a</sup>	46.87	-0.72	8.99	-0.30

<sup>a</sup> *S. cerevisiae* and *B. clausenii*

Yield (Y) with Genencor I, a mixed culture, and a *Populus* type substrate at a concentration other than 7.5% is calculated by ratioing the yields at the new cellulose concentration to the yield at a 7.5% cellulose



Table 3-9. SSF Performance Data for Different Substrate Concentrations

Substrate	Cellulose Concentration (g/L)	Cellulose Loading (IU/g)	Cellulase Type	Ethanol		Residence Time (d)	Organism
				Yield (%)	Concentration (g/L)		
Sigmacell 50	75	7	Genencor I	35.1	14.9	2	<i>S. cerevisiae</i>
"	100	"	"	34.2	19.4	"	"
"	150	"	"	29.7	25.2	"	"
"	75	13	"	45.7	19.1	"	"
"	100	"	"	44.1	25.0	"	"
"	150	"	"	38.7	32.9	"	"
"	75	26	"	56.6	24.1	"	"
"	100	"	"	50.4	28.5	"	"
"	150	"	"	43.2	36.7	"	"
"	75	7	"	49.5	21.8	4	"
"	100	"	"	50.4	28.5	"	"
"	150	"	"	41.4	35.2	"	"
"	75	13	"	61.1	25.1	"	"
"	100	"	"	56.7	32.1	"	"
"	150	"	"	51.3	43.6	"	"
"	75	26	"	70.1	29.8	"	"
"	100	"	"	62.1	35.2	"	"
"	150	"	"	53.1	45.1	"	"
"	75	7	"	69.2	29.4	4	"
"	100	"	"	65.7	37.2	"	"
"	150	"	"	57.6	48.9	"	"
"	75	13	"	78.2	33.2	"	"
"	100	"	"	72.9	42.3	"	"
"	150	"	"	68.4	58.1	"	"
"	75	26	"	81.1	34.8	"	"

Table 3-9. SSF Performance Data for Different Substrate Concentrations (Continued)

Substrate	Cellulose Concentration (g/L)	Cellulose Loading (IU/g)	Cellulase Type	Ethanol		Residence Time (d)	Organism
				Yield (%)	Concentration (g/L)		
Sigmacell 50	100	26	Genencor I	76.5	43.3	4	<i>S. cerevisiae</i>
"	150	"	"	71.1	60.4	"	"
"	75	7	"	46.7	19.9	2	Mixed I <sup>a</sup>
"	100	"	"	42.3	23.9	"	"
"	150	"	"	53.1	45.1	"	"
"	75	13	"	55.7	23.7	"	"
"	100	"	"	47.7	27.0	"	"
"	150	"	"	40.5	34.4	"	"
"	75	26	"	60.2	25.6	"	"
"	100	"	"	53.1	30.1	"	"
"	150	"	"	45.0	38.2	"	"
"	75	7	"	70.1	29.8	4	"
"	100	"	"	63.0	35.7	"	"
"	150	"	"	50.4	42.8	"	"
"	75	13	"	75.5	32.1	"	"
"	100	"	"	63.9	36.2	"	"
"	150	"	"	54.0	45.9	"	"
"	75	26	"	75.0	32.1	"	"
"	100	"	"	68.4	38.7	"	"
"	150	"	"	59.4	50.4	"	"
"	75	7	"	85.4	36.3	4	"
"	100	"	"	79.2	44.8	"	"
"	150	"	"	64.8	55.0	"	"
"	75	13	"	88.1	37.4	"	"
"	100	"	"	80.1	45.3	"	"

Table 3-9. SSF Performance Data for Different Substrate Concentrations (Concluded)

Substrate	Cellulose Concentration (g/L)	Cellulase Loading (IU/g)	Cellulase Type	Ethanol		Residence Time (d)	Organism
				Yield (%)	Concentration (g/L)		
Sigmacell 50	150	13	Genencor II	69.3	58.8	4"	Mixed <sup>a</sup>
"	75	26	"	88.1	37.4	"	"
"	100	"	"	83.7	47.4	"	"
"	150	"	"	70.2	59.6	"	"

Source: Spindler et al. 1988  
<sup>a</sup> *S. cerevisiae* and *B. clausenii*

concentration, both determined from the above equation for  $Y_{cc}$ . Then this ratio is multiplied by the yield at 7.5% cellulose using Genencor I, a mixed culture, and a *Populus* type substrate. The equation is given below:

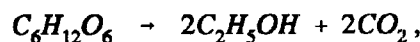
$$Y = Y_2 \left( \frac{Y_{cc}}{Y_{7.5}} \right),$$

where

$Y_{cc}$  = the yield at the new cellulose concentration (%), and  
 $Y_{7.5}$  = the yield at 7.5% cellulose (%).

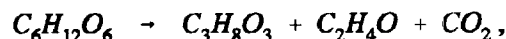
$Y_{cc}$  is calculated at the effective cellulose concentration in SSF (actual cellulose present plus potential cellulose that is present as ethanol). Because ethanol is inhibitory to the fermentation, its presence must be accounted for in the yield calculation. However, the above modification applies only to effective cellulose concentrations above 7.5%. Below this value no data are available and the yields at 7.5% are assumed to apply. This is a conservative assumption, as the yield may be higher.

The conversion of glucose to ethanol and production of carbon dioxide is governed by the following equation:



yielding 0.51 lb ethanol/lb glucose and 0.49 lb  $CO_2$ /lb glucose. Cellulose is converted to the equivalent weight of glucose by multiplying by 1.11, which accounts for the hydrolysis of cellulose to glucose.

It is assumed that 4.9 wt % (Chem Systems 1984) of the available potential glucose from cellulose is converted to glycerol and acetaldehyde according to the following equation:



yielding 0.51 lb glycerol/lb glucose, 0.244 lb acetaldehyde/lb glucose, and 0.244 lb  $CO_2$ /lb glucose. Cellulose is also converted to fusel oils (0.1 wt %, Chem Systems 1984) and 10.0 wt % is assumed converted to yeast cells (Spindler 1989a) with a yield of 0.5 lb cells/lb glucose, and the remaining cellulose is unconverted. With these assumed values, the highest ethanol yield that could be achieved is 85%. If the experimental data gives greater yields, these numbers would need to be reevaluated. For this analysis, SSF temperature is assumed to be 37°C, pH is uncontrolled, and the recycled water is assumed to supply all the required nutrients for the fermentations.

The agitation power for SSF is assumed to be 0.1 hp/1,000 gal. This assumption is based on technical information derived from visits to existing corn-to-ethanol plants and subcontract work by Elander (1988) at CSU. The heat of fermentation for SSF is estimated from the heat of reaction of cellulose to ethanol (256 Btu/lb cellulose) plus the agitation energy input (0.1 hp/1000 gal), which gave a heat duty of 0.67 MMBtu/h/fermenter for a total heat duty of 18.1 MMBtu/h. All other heat input or losses (i.e., evaporation and conduction) are ignored.

The yeast culture for SSF in this study is a mixed culture of *S. cerevisiae* D<sub>5</sub>A and *B. clausenii* Y1414. Therefore, two strains of seed fermenters are required. Each culture inoculum for the SSF experiments is a 10% (v/v) inoculum grown on 1.0% glucose (Spindler 1989). In this study, this procedure is also used for all seed fermentations except for the final seed fermenter. For this case, purchasing enough glucose would have been too expensive. Therefore, glucose for the final seed cultures

is obtained by diverting 10% of the incoming cellulose and cellulase streams to the final seed fermenters for each culture (total of 20%). The hydrolysis of cellulose by the cellulase produces the glucose necessary for yeast growth. It is assumed that the fermentations are long enough to convert enough cellulose to produce a 1.0% glucose concentration. The entire contents of both seed fermenters, including unreacted cellulose, is fed to the SSF reactors.

The fermentation and cycle times for the *S. cerevisiae* fermentation is 1.0 and 1.5 days, respectively, and for the *B. clausenii* fermentation, 1.5 and 2.5 days, respectively. The aeration requirement for all seed fermenters is assumed to be 0.2 vvm. Agitation power for the final seed fermenters is assumed to be 0.5 hp/1,000 gal, and 1.0 hp/1,000 gal for all other seed fermenters. Fermentation temperatures are 37°C, pH is uncontrolled, and nutrients are assumed provided by recycled water. The cell yield is assumed to be 0.5 g/g glucose and CO<sub>2</sub> production is ignored. The heat duty for the seed fermentation is estimated as the heat of combustion of the cellulose actually converted to cell mass and is 8.8 MMBtu/h.

## Section 4.0

### Process Design of the Base Case

#### 4.1 Design Basis

The design basis for the base case process design follows.

Plant type	Grass roots, N <sup>th</sup> plant
Plant location	Unspecified
Onstream time	8000 h/yr
Feed	1.0-in. wood chips
Nominal capacity	160,000 lb dry wood/h (1,920 ton/d)
Feed composition (dry basis)	46.2 wt % cellulose 24.0 wt % xylan 24.0 wt % lignin 5.6 wt % solubles 0.2 wt % ash
Feed moisture content	50.0 wt %
Product	Denatured fuel 90.3 wt % ethanol 4.7 wt % water 5.0 wt % gasoline
Ambient air temperature	20°C (68°F)
Pretreatment	Dilute sulfuric acid hydrolysis
Cellulose conversion	SSF
Cellulase production	<i>T. reesei</i>
Xylose conversion	Genetically engineered <i>E. coli</i>
Lignin utilization	Boiler fuel
Environmental	Soluble waste organics sent to anaerobic/ aerobic treatment  Solid waste organics (lignin, etc.) sent to boiler as fuel

	Cooling tower blowdown sent to evaporation pond
	Gypsum and boiler ash sent to off-site disposal
	Fermentative CO <sub>2</sub> vented to atmosphere
	Flume pond drained to evaporation pond
Utilities	
Steam	On-site generation from solid waste organics (lignin, etc.) and from methane and sludge from anaerobic/aerobic waste treatment
	Steam pressure levels: 50 psig (298°F) 150 psig (366°F)
Electricity	On-site generation from excess steam Excess electricity sold over the fence
Cooling water	
	Temperature 30°C (86°F) Maximum allowable rise 14°C (25°F)
Chilled water	
	Temperature 10°C (50°F) Maximum allowable rise 8°C (15°F)
Fermentation air	45 psig, 28°C
Plant air	100 psig
Cleaning and sterilization	Clean-in-place and chemical sterilization

## 4.2 Process Description

The following process descriptions describe the flow of materials between pieces of equipment and give all data and assumptions, not already given in Section 3.0, used to design the process. All the data and assumptions used to design the process are listed in a more concise form in Appendix B and the process flowsheets are given in Appendix E. Equipment numbers are given in the description for each piece of equipment on the flowsheets.

### 4.2.1 Wood Handling (Area 100)

Freshly cut 1.0-in. wood chips are delivered by large 23-ton trucks to receiving stations, which consist of a hydraulic truck dumper with scale (GY-101A/B/C/D). Chips are off-loaded into a washing flume pond that allows separation of heavy tramp material from the chips. Chips are swept by water flow from

the flume pump (PP-101) to be picked up by a radial stacking conveyor (GS-101), which transfers the chips to the wood pile. Periodically, the flume pond is emptied by pumping the water to the evaporation pond with the flume pump. Then any tramp material is removed from the bottom of the flume pond.

The wood pile holds a 4-day supply of chips and is managed on a first-in, first-out basis by front-end loaders (GM-101A/B/S). It is assumed that there is no degradation in the cellulose or xylan during this time. Chips are fed by a front-end loader to a belt conveyor (GS-102) that delivers the chips to the disk refiners (GG-101A/B/C/D) (see Appendix A-1 for a discussion of wood chip milling). Along the conveyor, small magnetic items down to 0.5 in. (12.7 mm) are removed by a magnetic chip cleaner (GS-103). The chips are milled down to 1/8 in. (3 mm). The milling power requirements are about 128 hp/dry ton. The material exiting the disk refiners drop directly into milled chip belt conveyor GS-102 and are delivered to the screw feeders (GS-202A/B).

#### **4.2.2 Pretreatment and Neutralization (Area 200)**

Screw feeders (GS-202A/B) delivers wood at a controlled rate to the impregnation reactors (MR-201A/B), which are continuous pulp digesters made of Carpenter 20 stainless steel. The reactors operate at atmospheric pressure. From the sulfuric acid storage tank (T-201), acid is pumped by the H<sub>2</sub>SO<sub>4</sub> pump (PP-201A/S) to a line mixer (GA-201) and mixed with hot (100°C) recycled water obtained from the bottom of the beer distillation column after separating out the solids. The resulting sulfuric acid solution is sprayed onto the wood in the impregnation reactors. Low-pressure steam (50 psig) is also added to the impregnation reactors to bring the wood slurry temperature up to 100°C. A residence time of 10 min ensures sufficient wetting of the biomass by the acid and water and complete migration of acid into the wood structure. The solids concentration after steam, water, and acid addition is 35%.

The wood slurry then drops into the prehydrolysis reactors (MR-202A/B), operating at 105 psig through a high-pressure rotary valve. These reactors are continuous pulp digesters made of Carpenter 20 stainless steel. High-pressure saturated steam (150 psig) is then added to bring the wood slurry temperature up to 160°C. The acid concentration after steam addition is 0.85 wt %. The residence time in the prehydrolysis reactor is 10 min, during which the xylan is converted as follows: 80.0% to xylose and 13.0% to furfural; the balance is unconverted. Cellulose is converted as follows: 3.0% to glucose and 0.1% to HMF; the balance is unconverted. The appropriate polymerization factors are 1.11 lb glucose/lb cellulose, 1.14 lb xylose/lb xylan, 0.78 lb HMF/lb cellulose, and 0.73 lb furfural/lb xylan.

After prehydrolysis, the slurry is discharged into a blowdown tank (T-203) operating at atmospheric pressure. Flash vapors are sent to the distillation section to preheat incoming feed to the beer column. It is assumed that all the furfural escapes in the overhead flash vapors. The slurry is mixed with hot (100°C) recycled water in the bottom half of the blowdown tank to achieve a 12 wt % solids concentration. It assumed that the residence time is 5.0 min and that the required mixing power is 2.0 hp/1,000 gal. The 12.0 wt % slurry is then pumped by hydrolyzate pump (PP-202A/S) to the neutralization vessel (T-206) to be mixed with lime.

Lime is delivered by rail car and unloaded into the lime unloading pit (MF-224). It is then transferred by the lime unloading conveyor (GS-223) to the lime storage bin (MB-220). The bin is vented to the atmosphere through the lime dust cyclone (GC-227). Lime is added directly to the neutralization vessel by the lime solids feeder (GS-225) and the reaction stoichiometry and solubility are as described in Section 3.2 and Appendix C-6. It is assumed that the residence time is 10 min and that the required mixing power is 2.0 hp/1,000 gal. The neutralized slurry is then pumped by the neutralized hydrolyzate pump (PP-203A/S) through the feed cooler (TT-220), using water to cool the slurry to 37°C. cooling water. The temperature rise of the cooling water is 14°C. The flow out of the cooler is then split into two streams; one is sent to xylose fermentation and the other is sent to cellulase production.



The fraction of the cellulose stream from the neutralization vessel split off to cellulase production is determined from a balance of cellulase consumption in SSF versus cellulase production as follows:

$$XLm_c = (1-X)(m_c + m_x)Y,$$

where

- X = fraction of stream to SSF,
- L = cellulase loading in SSF (IU/g cellulose),
- m<sub>c</sub> = cellulase flow rate (lb/h),
- m<sub>x</sub> = xylose flow rate (lb/h),
- Y = cellulase yield (IU/g cellulose).

### 4.2.3 Xylose Fermentation (Area 300)

Most of the xylose-containing wood slurry from the neutralization tank (T-206) is pumped to the xylose fermentation area. In turn, a small fraction of this material is split out to the xylose seed fermenter, with the bulk going to the main xylose fermenters.

Xylose fermentation is anaerobic and takes place in eight continuous stirred tanks (750,000 gal) connected in series. Fresh inoculum (10% v/v) and the wood slurry are combined continuously in the first fermenter (FM-303A). The material then flows to each subsequent fermenter because of a hydraulic head difference between the first and last fermenters. Each fermenter is at atmospheric pressure and filled to 95% of its total volume, pH is controlled at 6.5, and temperature is maintained at 37°C by cooling water flowing through internal coils. The cooling water rise is 3°F and the total heat duty for all the fermenters is 14.6 MMBtu/h. The ethanol yield is approximately 95% for a residence time of 2 days and base usage is taken from Table 3-2. It is assumed that only 90.0% of the sugars (both xylose and glucose) would be available for conversion to ethanol because of sugars trapped within the wood particles. This reduces the overall ethanol yield to 85.5% (0.95 × 0.90). It is also assumed that the mixing power requirements are 0.1 hp/1,000 gal and that the recycled water contains all the nutrients necessary to support microorganism growth. Ammonia for pH control is stored in the base tank (T-321) and used at a rate of 0.2583 lb/h of ethanol produced. CO<sub>2</sub> production and microorganism growth are calculated as described in Section 3.3.

*E. coli* inoculum is grown in a series of six batch seed fermenters (FM-305 through 310), filled to 95% of their capacity on a combined 2.0 wt % xylose and glucose substrate. The fermentation time is 12 h/fermenter with an additional 12 h required for seed fermenter cleaning and turnaround. The temperature is 37°C and the pH is uncontrolled. Nutrients are assumed to be supplied by the recycle water, and CO<sub>2</sub> production is ignored. The cell yield is assumed to be 0.5 g cells/g sugars and a 10% (v/v) inoculum is used. It is assumed that the air flow rate to the seed fermenters is 0.2 vvm and that the mixing power is 0.1 hp/1,000 gal. The seed fermenters are cooled by internal coils with an allowable temperature rise of 3°F and the heat duty is 4.8 MMBtu/h.

The contents of the last seed fermenter flows by a pressure difference (supplied by the air compressors) to the seed hold tank (T-301), which is sized the same as the final seed fermenter, from which inoculum is delivered continuously to the first xylose fermenter by a pressure difference. The maximum mixing power requirement is also assumed to be 0.1 hp/1,000 gal in the seed hold tank and the average power required is assumed to be one-half this value because the tanks would be on average only half full.

Off gas from each of the xylose fermenters passes through an exhaust condenser (TT-312), which is cooled by chilled water to remove water and ethanol from the off gas. The chilled water temperature rise is 2°C. The amount of ethanol leaving the fermenters is determined by an Aspen simulation using a flash calculation and assuming a linear increase in ethanol concentration in each fermenter. Similarly, the ethanol condensed in the exhaust condenser is also determined from an Aspen simulation. The remaining carbon dioxide is vented and condensed water and ethanol are collected and sent to the ethanol vent storage tank (T-507).

#### 4.2.4 Cellulase Production (Area 400)

The xylose-containing wood slurry from the neutralization tank (T-206) is pumped to the sterile feed tank (T-405), which is sized for a 2-day residence time, and is held until needed by the cellulase production or seed fermenters. The maximum mixing power requirement for the sterile feed tank is assumed to be 2.0 hp/1,000 gal and the average power consumption is one-half this value.

Feed is delivered to the aerobic batch cellulase fermenters (FM-401A/B/C) and the inoculum to each fermenter is 5% (v/v). There are three 250,000-gal fermenters, each having a fermentation time of 5.7 days and a total cycle time of 6 days. The working capacity of each fermenter is 80%. The fermenters are operated in a staggered fashion with a new fermentation beginning every 2 days. Corn steep liquor and nutrients are sterilized (121°C, for 15 min) in the media prep tank (T-400) by external steam heating and then pumped by the prep tank transfer pump (PP-411A/S) to the cellulase fermenter. Recycled water at 100°C is cooled by chilled water (temperature rise of 8°C) in the water cooler (TT-402) to 28°C and added to a fermenter. Producing a final cellulose/xylose concentration of 5% in the fermenter and the nutrient concentrations shown in Table 3-4. The fermentation is conducted at a temperature of 28°C, at a pH of 4.8 and a head pressure of 10 psig.

The average cellulase production rate ( $m_e$ , lb/h) is calculated from the carbohydrate (cellulose and xylose) flow rate ( $m_c$ , lb/h) to the cellulase production fermenters, the average cellulase yield shown in Table 3-3 ( $Y$ , IU/g of carbohydrate where carbohydrate is equivalent to cellulose in Table 3-3), and the specific activity ( $S_a$ , IU/g enzyme, value shown in Table A-4) according to the following formula:

$$m_e = m_c \left( \frac{Y}{S_a} \right).$$

It is assumed that all of the cellulose and xylose sent to the cellulase production fermenters is consumed. The inert components—lignin, soluble solids, and ash—are assumed unchanged. Cell mass is calculated from the fermenter working volume, assuming a final cell concentration of 20 g/L. CO<sub>2</sub> production is ignored as described in Section 3.4. The oxygen uptake rate is 42 mM O<sub>2</sub>/l-h, the air flow rate is 0.15 vvm, the mixing power is 1.18 hp/1,000 gal, and the heat duty is 13.1 MMBtu/h. The fermenters are cooled with chilled water using internal coils. The chilled water temperature rise is 8°C. Off gas from the fermentation is passed through the fermenter exhaust condenser (TT-422) (8°C temperature rise) to remove water vapor, which is cooled by chilled water, and then is vented to the low-pressure vent system.

Ammonia is added during the fermentation as needed from the base storage tank (T-321) at a rate of 0.045 lb/lb of cellulose and xylose. The base storage tank is sized at 1% of cellulase fermentation capacity. Antifoam (corn oil) is sterilized (121°C, for 15 min) in the antifoam tanks (T-403A/B) by direct steam injection and transported by a pressure difference as needed to the cellulase fermenter. The total antifoam usage is 1.0 mL/L of fermenter volume (A.D. Little 1984).

*T. reesei* inoculum is obtained from a series of atmospheric batch aerobic seed fermenters (FM-401A/B through 404A/B) filled to 80% capacity and transported by pressure difference (supplied by air compressors) to the cellulase fermenter. The fermentation time for each seed fermenter is 3.5 days and the total cycle time is 4.0 days. The inoculum for all seed fermenters is 5% (v/v). For the last seed fermenter, the initial substrate and media is assumed to be a 1.0 wt % combined cellulose and xylose with the media composition shown in Table 3-4. For all other seed fermenters, 1% glucose with the same media composition is assumed. Cooled recycle water is used for dilution purposes. The temperature of the seed fermentations is 28°C and the pH is assumed controlled to 4.8. It is assumed that no base is used in the seed fermenters, as discussed in Section 3.4. The cell yield is assumed to be 0.5 g cells/g cellulose and xylose.

It is assumed that the air flow rate to each seed fermenter is 0.2 vvm. Fermentation air is sterilized by a 2.0- $\mu$ m filter and injected into the bottom of each fermenter. The mixing power requirement is 0.5 hp/1,000 gal for the final seed fermenter and 1.0 hp/1,000 gal for all other seed fermenters. Each seed fermenter is cooled by chilled water passing through internal coils with an allowable temperature rise of 8°C. The heat duty is 0.66 MMBtu/h.

At the end of the cellulase fermentation, the broth is pumped (PP-403A/S) to the cellulase hold tanks (T-410A/B) and then pumped continuously from the hold tank by the cellulase feed pump (PP-412A/S) to the first SSF fermenter (FM-500A). Two hold tanks are used so that each tank can be completely emptied before being refilled, thus eliminating any holdup that would result in loss of enzyme activity. Each hold tank is the same size and the maximum mixing power requirement is the same as the cellulase production fermenters. It is assumed that the average power usage by the hold tank agitator is one-half the maximum.

#### **4.2.5 Simultaneous Saccharification and Fermentation (Area 500)**

The stream from the last xylose fermenter (FM-303H) flows by hydraulic head differential to the SSF fermentation area. Twenty percent of the stream is split off to the SSF seed fermenters and the rest is sent to the first SSF fermenter (FM-500A), where it is combined with cellulase from the cellulase hold tank (T-410A/B) (at a cellulase loading of 7 IU/g cellulose) and *S. cerevisiae* inoculum (10% v/v) and *B. clausenii* (10% v/v) inoculum. The SSF reactor system consists of 27 750,000-gal continuous anaerobic stirred tanks at atmospheric pressure and connected in series. Flow between fermenters is by a hydraulic head differential. Each fermenter is filled to 95% of its capacity, and the total residence time in all 27 fermenters is 7 days. The temperature is maintained at 37°C, and the pH is not controlled. No extra nutrients are provided. Cellulose yield is calculated from the correlations described in Section 3.5. In addition, 0.1% of the cellulose goes to fusel oils (Chem Systems 1984), 4.9% to glycerol/acetaldehyde (Chem Systems 1984), and 10% to cell mass (Spindler 1988). It is assumed that 80% of the xylan left unconverted by the pretreatment process is converted to xylose by the xylanases contained in the cellulase broth. The cell yield is assumed to be 0.5 g cells/g cellulose. CO<sub>2</sub> production is as described in Section 3.5. The mixing power requirement is 0.1 hp/1,000 gal. The total heat duty is 18.1 MMBtu/h and cooling is accomplished by cooling water in internal coils with a 3°C allowable temperature rise.

The yeast inoculums (*S. cerevisiae* and *B. clausenii*) are obtained from a series of six batch aerobic seed fermenters (FM-501A/B through 506A/B), each filled to 95% capacity. The fermentation time is 1 day and 2 days for *S. cerevisiae* and *B. clausenii*, respectively, and total cycle time is 1.5 days and 2.5 days, respectively. The inoculum for each subsequent fermentation is 10% (v/v). The substrate for the seed fermenters is purchased 1.0 wt % glucose, except for the final seed fermenters, which use a portion of the wood slurry and cellulase stream to produce the 1.0 wt % glucose. The cellulase loading is the same as the main SSF fermenters (7 IU/g cellulose). The temperature is 37°C, the pH is uncontrolled, a cell yield of 0.5 g cell/g glucose is assumed, and CO<sub>2</sub> production is ignored.

It is assumed that the air flow rate is 0.2 vvm and that the mixing power requirements are 0.5 hp/1,000 gal for the final seed fermentation and 1.0 hp/1,000 gal for all other seed fermentations. The heat duty is 8.8 MMBtu/h and the seed fermenters are cooled with cooling water flowing through internal coils with an allowable temperature rise of 3°C. The contents of the final seed fermenters are transported by pressure difference to the seed hold tanks (T-501A/B), which are sized the same as their respective seed fermenters, and then transferred continuously by pressure difference to the first SSF fermenter. The maximum mixing power requirement for the seed hold tanks is assumed to be 0.1 hp/1,000 gal, and the average power requirement is assumed to be one-half the maximum.

Off gas flows through the exhaust condenser (TT-525), which uses chilled water with a 2°C temperature rise, to remove water and ethanol vapor from the gas stream. The composition of the vent and condensate streams from the exhaust condenser is determined by an Aspen simulation. The condensate, which is enriched in ethanol, is combined with the condensed ethanol stream from the xylose fermenters and stored in the ethanol vent storage tank (T-507). The contents of this tank are pumped by the ethanol vent pump (PP-509) to the rectification column (AS-602) to further remove water. The off gas, containing mainly carbon dioxide, is vented.

#### **4.2.6 Ethanol Purification and Solid Separation (Area 600)**

Dilute beer from the last SSF fermenter (FM-500AA) is pumped by the beer transfer pump (PP-505A/S) through the feed cross exchanger (TT-615) to be heated to 64°C by exchange with flash vapor (100°C) from the blowdown tank (T-203). The dilute beer is further heated by 50 psig condensing steam (298°C) in the feed preheater (TT-613). The rest of the design of the distillation system is by Badger Engineers, Inc. (1984), except that heat exchanger calculations were performed by SERI staff members. Then beer, along with the bottom steam from the rectification column (AS-602), enters the degasser drum (T-601), where some of the dissolved gases (primarily CO<sub>2</sub>) leave as flash vapor. The degasser drum condenser (TT-602) uses cooling water with an allowable 25°F temperature rise to condense ethanol and water from the vent gases. Uncondensed vapor passes on to the low-pressure vent system. Degassed beer then enters the beer column (AS-601). Heat to the beer column is supplied through a thermosiphon reboiler (TT-603), using low-pressure steam. The bottoms are pumped by the beer column bottoms pump (PP-601A/S) to lignin separation.

The overheads from the beer column (containing 40 wt % ethanol, Badger 1984) are condensed by the beer column condenser (TT-605), using cooling water (25°F temperature rise), and are collected in the beer column reflux drum (T-602). The beer drum vent condenser (TT-606) uses cooling water (25°F temperature rise) to condense vaporized ethanol and water and return them back to the reflux drum. Remaining vapors pass on to the low-pressure vent system. The beer column reflux pump (PP-603A/S) returns a portion of the liquid from the reflux drum back to the beer column and sends the rest to the rectification column (AS-602). The reflux ratio for the beer column is 0.4 (Badger 1984).

In addition to the feed from the beer column, the rectification column also receives the condensed ethanol stream from the ethanol vent storage tank (T-507). This is added near the top of the column. The bottom stream from the rectification column is pumped by the rectification column bottoms pump (PP-607A/S) to the degasser drum (T-601). Heat is supplied to the rectification column through the thermosiphon reboiler (TT-609), using low-pressure steam as the heat source. The overhead vapor is condensed with cooling water (25°F temperature rise) in the rectification column condenser (TT-610) and collected in the rectification column reflux drum (T-605). Vapors from the reflux drum are condensed with cooling water (25°F temperature rise) in the rectification column vent condenser (TT-611) and returned to the reflux drum. Remaining vapors pass on to the low-pressure vent system. A portion of the liquid from the reflux drum is pumped by the rectification column reflux pump (PP-608A/S) back to

the rectification column. The rectification column reflux ratio is 1.6 (Badger 1984). The rest of the liquid, containing 95 wt % ethanol, is pumped to the ethanol product tank (T-701A/B).

Fusel oils, chiefly amyl, isoamyl and propyl alcohol, will accumulate within the rectification column if not removed. Therefore, a small side stream is withdrawn from the rectification column, mixed with process water (5.0 lb/lb fusel oils, Badger 1984) and cooled with cooling water (25°F temperature rise) in the fusel oil cooler (TT-607). This stream then enters the fusel oil decanter (T-603). There the fusel oil is decanted and pumped by the fusel oil pump (PP-605A/S) to the boiler (HB-901). Water and ethanol are returned to the rectification column by the wash return pump (PP-604A/S).

The stream from the bottom of the beer column is combined with sump solids from the recycled water tank (T-630) and sent to the centrifuge (GC-609A/B/C). The centrifuge is assumed to recover 95% of the solids and to concentrate the solids to 50 wt %. Concentrated solids are transported by sludge screws (GS-611A/B) to the boiler (HB-901). Water overflow from the centrifuge, which still contains a small fraction of solids, is sent to the recycled water tank. Solids collecting on the bottom of the tank are sent back to the centrifuge by the sump pump (PP-632). A recycled water pump (PP-631A/S) delivers water to waste treatment or mixes recycled water with fresh process water and recycles it back to the process. Fresh process water is added to reduce the level of organics buildup in the recycled stream.

#### **4.2.7 Off-site Tanks (Area 700)**

Ethanol from the rectification reflux drum (T-605) is pumped (PP-608A/S) to the ethanol product tanks (T-701A/B) (sized for 10 days' storage). Prior to entering the tank, 95 lb of ethanol product is blended with 5 lb of gasoline. Gasoline is delivered by truck or rail and stored in the gasoline storage tank (T-710) (sized the same as Badger 1984). Gasoline is blended with ethanol by using the gasoline-blending pump PP-710A/S. Denatured fuel ethanol is then pumped to the shipping point by the ethanol export pump (PP-701A/B/S).

Corn steep liquor (CSL) is delivered to the CSL tank (T-720) (sized for 30 days' storage) by truck or railcar. The CSL transfer pump (PP-720A/S) delivers CSL to the media prep tank (T-400) as needed.

Concentrated sulfuric acid is delivered by truck or railcar to the sulfuric acid storage tank (T-703) (sized for 9 days' storage based on the Badger [1984] design). Moisture is kept from the acid by a desiccant air filter (GF-703). A sulfuric acid transfer pump (PP-703A/S) transports acid to the H<sub>2</sub>SO<sub>4</sub> tank (T-201).

Liquid ammonia is delivered by truck or railcar to NH<sub>3</sub> storage tanks (T-706A/B) (sized for 4 days' storage). Liquid ammonia is pumped as needed by the NH<sub>3</sub> transfer pump (PP-706A/S) to the base tank (T-321).

Corn oil, which is used as an antifoam agent, is delivered by truck or railcar to an antifoam storage tank (T-707) (sized for 3 weeks' storage). Corn oil is pumped as needed by the antifoam transfer pump (PP-707A/S) to the antifoam tank (T-403A/B).

Diesel fuel is delivered by truck and stored in the diesel fuel tank (T-708) (sized the same as Badger [1984]). Fuel is delivered to mobile equipment by the diesel fuel pump (PP-708A/S).

Fire water is stored in the fire water tank (T-704) (sized the same as Badger 1984) and delivered to the fire water mains with the fire water pump (PP-704A/S).

#### **4.2.8 Waste Treatment (Area 800)**

The waste treatment system consists of three sections: anaerobic digestion, aerobic digestion, and a low-pressure vent system. The design of this system was taken from the study of Badger Engineers, Inc. (1984). All design data were from Badger (1984) except the methane production rate, which was obtained from C. Rivard of SERI (1990). Because the anaerobic process is proprietary to Badger and no design information is given, some assumptions are made in this section of the plant. Wastewater from lignin separation (PP-631A/S), the clean-in-place system (PP-960), the feed cross exchanger (TT-615), and the low-pressure vents (MS-810) flow to and are blended in the equalization tank (T-803). Reactor feed pump (PP-808A/S) pumps hot wastewater to the feed cooler (TT-802), which lowers the temperature to 35°C by exchange with cooling water (25°F temperature rise). The wastewater is mixed with a portion of the anaerobic reactor effluent and fed to the anaerobic reactor (T-804), which produces methane. The anaerobic reactor converts 90% of the effluent organics (lignin is the only organic not converted), of which 0.8 lb of biogas are produced/lb of organics converted. The balance is assumed converted to cell mass. The biogas is 70.0 mol % methane and 30.0 mol % carbon dioxide (Rivard 1990). The gas passes through the off gas suction knockout pot (MS-806) to the off gas blower (PB-810). The off gas is compressed and sent to the off gas knockout drum (MS-809). System pressure is maintained by recycling off gas back to the outlet of the anaerobic reactor through the off gas cooler (TT-801) (size assumed to be 500 ft<sup>2</sup>), which cools with cooling water, or by using makeup nitrogen. Excess system pressure will divert off gas to the off gas burner (GO-806) to be flared. All gas produced, except that which is flared, is sent to the boiler as fuel. Liquid effluent from the anaerobic reactor (T-804) overflows into the reactor surge drum (MS-805) and is pumped by the reactor recycle pump (PP-809) to the biotreater (T-807). A portion is recycled back to the anaerobic reactor.

The remaining wastewater chemical oxygen demand (COD) in the liquid effluent from the anaerobic digester is removed by the aerobic digestion system. The wastewater from the anaerobic reactor is combined with overflow from the sludge centrifuge (GC-801) and fed to the biotreater (T-807), where all remaining organics are degraded except lignin. Air is injected through the biotreater aerators (GV-807) into the biotreater by the aeration blower (PB-812A/S). Effluent from the biotreater is combined with the sand filter backwash and boiler blowdown wastewater and is fed to the secondary clarifier (GV-808). Overflow from the clarifier is pumped by the final effluent pumps (PP-816A/S) to the process water tank (T-901).

The underflow from the clarifier is pumped by the sludge pump (PP-813A/S) to the sludge centrifuge (GC-801). Overflow from the centrifuge is sent back to the biotreater and underflow, assumed to contain 100% of the solids at a solids concentration of 50%, is fed to the sludge screw (GS-801) and delivered to the boiler (HB-901).

All low-pressure vents are drawn into the low-pressure vent knockout drum (MS-810) where entrained liquids are removed and sent to the anaerobic reactor (T-803). The remaining gas is transported to the boilers (HB-901A/B) by the low-pressure vent blower (PB-817A/S). A bypass around the blower is used for pressure control.

#### **4.2.9 Utilities (Area 900)**

**4.2.9.1 Boiler and Steam Distribution (BESSF-910).** The boiler (HB-901) is designed to burn liquid, gaseous, and solid fuels and to generate 1100 psia steam with 300°F of superheat. The boiler is sized to handle the waste streams from the plant. Gaseous and liquid fuels are burned directly and wet solids are first sent to a drying system that dries and fluidizes the solids into the boiler using boiler flue gas. Based on vendor calculations, the boiler efficiency, including drying, is 83.6%. Ash and gypsum solids left over after combustion are sent to off-site disposal. Electrical power is generated by letting steam down through

the turbogenerator. Steam is let down to the two levels required by the process, 150 psig (366°F) and 50 psig (298°F), and any remaining steam is condensed at 89 mm Hg. Condensate is returned to the boiler feed water (BFW) system and recycled back to the boiler.

**4.2.9.2 Boiler Feed Water System (BESSF-930).** This system is based on the design of Badger (1984) and is sized by ratioing our flow rate to the flow rate of the Badger design. All recoverable condensate from the steam system is collected in the condensate collection tank (T-930) and then pumped by the condensate pump (PP-910A/S) through the condensate polisher (GU-904A/S) to the condensate surge drum (MS-904). Fresh makeup water is added to the condensate surge drum through the demineralizer (GU-903A/B). The fresh water makeup rate is assumed to be 3.0% of steam usage plus steam used for direct injection into the process (i.e., steam injected into the wood slurry during impregnation and prehydrolysis) and steam lost in the deaerator. Condensate and makeup water is transferred by the deaerator pump (PP-909A/S) to the deaerator (GV-906). The deaerator operates at 10 psig (68.9 kPa) and expels air and steam to the atmosphere. Low-pressure steam for deaeration is supplied to the deaerator from the 50 psig distribution header, by flashes from the condensate collection tank, and from the boiler blowdown flash drum (MS-902). Boiler blowdown is collected in the blowdown flash drum (MS-902) and then is pumped by the blowdown pump (PP-906A/S) to aerobic digestion.

Deaerated boiler feedwater is treated with hydrazine and ammonia in the deaerator. Hydrazine is stored in the hydrazine drum (MS-903) and is pumped by the hydrazine transfer pump (PP-907) to be mixed with condensate in hydrazine addition unit (GU-907), which is then pumped to the deaerator. Ammonia is mixed with condensate in ammonia addition unit (GU-908) and then is fed to the deaerator. Phosphate dumped from bags is mixed with condensate in the phosphate addition unit (GU-909) and then used to dose the boiler steam drums. Deaerated and treated water is transferred by the high-pressure BFW pump (PP-908A/S) to the boilers.

**4.2.9.3 Process Water System (BESSF-920).** This system was based on the design of Badger (1984) and was sized by ratioing our flow rate to the flow rate of the Badger design. Process water is prepared by pumping well water with the well water pump (PP-913A/S) through a sand and anthracite filter (GF-901). A small fraction of the water (before filtration) is diverted to the wood washing pond. Process water transfer pump (PP-902A/S) feeds filtered water to the process water tank (T-901). Backwash feed pump (PP-904A/B) provides water for backwashing of the filter (GF-901). Backwash overflow from the filter is collected in the backwash transfer tank (T-905) and then transported by the backwash transfer pump (PP-905A/S) to the secondary clarifier (GV-808). Process water is distributed to the process water ring main by the process water circulating pump (PP-903A/S).

**4.2.9.4 Turbogenerator (BESSF-940).** The turbogenerator is sized based on the flow rate of steam from the boiler by ABB Sprout-Bauer (vendor quote). Based on that quote, it has an efficiency of 78.5%. The turbogenerator (GZ-911) is rated for 1100 psia, 300°F superheated steam. The unit includes a condenser, vacuum ejector set, and controls. The turbine condensate pump (PP-901A/S) returns condensate to the BFW system. The generator output is 13500 VAC, which is transformed to 480 VAC, 60 Hz, 3 phase, and 120/200 VAC, 60 Hz, 3 phase.

**4.2.9.5 Cooling Water System (BESSF-940).** The cooling tower is sized based on the required cooling water flow rate. The cooling tower (GT-912) provides cooling water at 30°C with a maximum allowable temperature rise of 14°C (25°F). Pressure in the cooling water distribution system is maintained at 60 psig (410 kPa) by the cooling water pumps (PP-912A-F/S). Cooling tower blowdown containing dissolved salts is discharged directly to an evaporation pond. The water losses are 1.3% of flow for evaporation, 0.3% of flow for windage, and 2.7% of flow for blowdown (Walas 1988).

**4.2.9.6 Plant and Instrument Air Systems (BESSF-940).** Plant and instrument air are provided by the air compressor (PC-910) (100 psig). This compressor charges the plant air receiver (MS-906), from which plant air is withdrawn. The instrument air receiver (MS-907) receives air from the plant air receiver through the instrument air dryer (GY-910), from which instrument air is withdrawn.

**4.2.9.7 Sterile Air System (BESSF-950).** Sterile air is provided by an air compressor package (PK-950A/B/S), which is sized based on the required air flow rate by Ingersoll-Rand (vendor quote). The air is cooled to 28°C by intercoolers and then sent through a prefilter to the fermentation air users.

**4.2.9.8 Chilled Water System (BESSF-950).** Chilled water is provided at 10°C (50°F) by the chilled water package (PK-950), which includes circulation pumps. This package is sized based on the required chilled water flow rate. The maximum allowable temperature rise is 8°C (15°F).

**4.2.9.9 Clean-in-Place and Chemical Sterilization (CIP/CS) System (BESSF-950 and BESSF-960).** The clean-in-place and chemical sterilization system is designed to clean and sterilize fermenters and tanks prior to use. Sterile water used by the system is first obtained by heating process water in the water sterilizer (TT-953) with low-pressure steam. This water is held for 15 min in a coil and then stored in the sterile water tank (T-953). On demand, sterile water is fed to the CIP/CS system by the sterile water pump (PP-953).

Cleaning chemicals are mixed with sterile water in the cleaning tank (T-961). The solution is pumped to the appropriate vessel by the supply pump (PP-960A/S). The cleaning solution is collected and returned to the cleaning tank by the CIP/CS sump pump (PP-965A/B/C/S). A separate sump pump is used for each area. Then the vessel is rinsed with sterile water and this water is collected by the CIP/CS sump pump and sent to waste treatment. Next, sterilizing solution from the sterilization tank (T-960) is pumped (PP-960A/S) to the appropriate vessel and collected by the CIP/CS sump pump and returned to the sterilization tank. Finally, the vessel is rinsed with sterile water from the sterile rinse water tank (T-963), which is also collected by the CIP/CS sump pump and returned back to the sterile rinse water tank. Periodically, the contents of the sterile rinse water tank (T-963), the sterilization tank (T-960), and the cleaning tank (T-961) are dumped to the waste treatment system by the supply pump (PP-960A/S).

No design information was available for the CIP/CS system, so it was arbitrarily sized at 10,000-gal tanks and for an average flow rate of 320 lb/h.

### **4.3 Process Flowsheets and Material Balance**

This section presents the detailed process flowsheets for the overall biomass-to-ethanol process. Each flowsheet includes equipment configuration and type and material balance information. Process descriptions for these flowsheets were given in the preceding section.

### **4.4 Process Utility Summary**

The utility summary for the base case plant is shown in Table 4-1. Steam requirements are determined from the material balance calculations. Steam is generated in the boiler from the burning of lignin, unreacted solids, and methane produced from the anaerobic digestion of the waste stream. This steam is let down to the two levels required by the process, 150 psig (366°F) and 50 psig (298°F). Low-pressure steam (50 psig) is used primarily in the reboilers of the beer and rectification columns in the ethanol purification area of the plant. The next major steam use is for preheating the wood slurry in the impregnation reactor during pretreatment. High-pressure steam (150 psig) is required only during pretreatment to obtain the high temperature (160°C) needed for prehydrolysis.



Table 4-1. Utilities Summary

Area	Steam		Electricity Consumed (MM Btu/h)	Cooling Water (gpm)	Chilled Water (gpm)	Fermentation Air (lb/h)
	50 psig (lb/h)	150 psig (lb/h)				
100 Wood handling			26.0	400		
200 Pretreatment	30,597	41,354	1.0	7,937		
300 Xylose fermentation			2.1	13,446	55	40,013
400 Cellulase production	122		2.5		1,774	48,288
500 SSF			7.8	16,100	158	159,150
600 Ethanol purification	171,986		1.1	10,467		
700 Off-site tankage			0.3			
800 Waste treatment			1.3	3,619		
900 Utilities	1064		34.7	15,467	1,392	
Miscellaneous	20,000		2.0			
Total	222,769	41,354	78.8	67,436	3,379	247,451

Well water requirements: 2977 gpm

Steam to turbogenerator

Steam condensed	169,671 lb/h
Steam to 150 psig	41,354 lb/h
Steam to 50 psig	222,853 lb/h
Total	433,878 lb/h

Electricity produced	123.2 MM Btu/h (36.1 MW)
Electricity consumed	78.8 MM Btu/h (23.1 MW)
Electricity exported	44.4 MM Btu/h (13.0 MW)

Electricity usage (23.1 MW) is taken from the equipment specifications shown in the equipment list of Section 4.5. Electricity is produced (36.1 MW) by letting high-pressure steam from the boiler down through a turbogenerator. All plant electrical requirements are met and the additional power (13.0 MW) is sold to the grid. At the bottom of Table 4-1 is the total steam production from the boiler and its distribution to various pressures, total power generated, electricity consumption, and export. The utilities area is the largest user of electricity because of the air compressors and chilled water system. Within the process, the largest single user of electricity are the disc refiners in wood handling area, which is based on an ABB Sprout-Bauer vendor quote. The SSF area is also a large user of electricity because of the agitation requirements for the fermenters.

Cooling water rates are determined from material balance information, stream temperatures and enthalpies, and assumed cooling water rises. The temperature of the cooling water is 30°C with a maximum allowable temperature rise of 14°C. Cooling water is used extensively in xylose fermentation and SSF for fermenter cooling, in the beer and rectification column condensers in ethanol purification, and in the utilities area for the turbogenerator condensers.

As with cooling water, chilled water rates are determined from material balance information, stream temperatures and enthalpies, and the assumed chilled-water temperature rises. The temperature of the chilled water is 10°C with a maximum 15°C allowable temperature rise. Chilled water is primarily required in cellulase fermentation for fermenter cooling and for interstage cooling of the air compressors in the utilities area.

Well water, as listed at the bottom of the table, is required to replace losses from the cooling towers and water injected into the process.

Fermentation air requirements are based on fermenter volumes and air rates as previously discussed. The largest user of air is the SSF seed fermenters, because of the large seed fermenter volumes.

## 4.5 Equipment List

This section presents the equipment list detailed by equipment type. The equipment types are pumps, solids handling equipment, heat exchangers, vessels, towers, and miscellaneous equipment. Each list contains specifications, the number required, the cost, and the cost source. Equipment is sized, either by SERI staff members or by vendors, based on the material balance information presented on the flowsheets and specifications presented in Appendix B and discussed in Sections 3.0 and 4.0, or by ratioing to sizes of equipment in the Badger study (1984), as previously discussed.

Pumps are sized based on the required flow rate and an estimate of the total head as given on the equipment list. From this information, the brake horsepower is determined and the required motor size is calculated from a motor efficiency curve (Ulrich 1984). Heat exchangers are sized from the flow rates and temperatures shown on the flowsheets and by using the heat transfer coefficients given in Appendix C-5. Vessel sizing is based on the flow rates, fill capacity, and residence times (given in the discussions in Sections 3.0 and 4.0). Agitator size (brake hp requirement) is based on the specifications given in Sections 3.0 and 4.0 (hp/1,000 gal), where the motor efficiency curve is again used to determine motor size. Solids handling equipment is sized from the information contained in Weber (1988), Ulrich (1984), and Perry and Chilton (1973).

The cost shown in the equipment list is in first-quarter 1990 dollars updated from the original cost source or from vendor quotes from 1 to 6 years ago. These are updated to 1990 dollars with the chemical engineering plant cost indexes published in *Chemical Engineering*. Costs for the equipment are developed from published cost curves, cost curves developed from previous ICARUS cost runs, and cost-estimating

software (Chemcost). Costs for certain critical pieces of equipment were obtained or verified by vendor quotes, because they represent a large fraction of the total cost of the plant. These items are the boiler, the turbogenerator, fermentation air compressors, SSF and xylose fermenters, ethanol product tanks, impregnator and prehydrolysis reactors, the screw feeder to the impregnator, and disk refiners. These items represent approximately 70% of the total base equipment cost.

#### 4.6 Plot Plan

A plot plan for the biomass-to-ethanol plant is presented in Drawing No. BESSF-1000. This is a preliminary layout based on equipment sizes when known and assumptions when unknown. Much of the layout used plot plan information from Badger (1984), which was for a plant having the same wood feed rate. The following buildings and areas are sized according to the Badger study; the maintenance building, the warehouse building, the administration building, the utilities area, and the off-site tanks.

The wood chip pile is sized based on the dimensions of the radial stacking conveyor and the angle of repose of the chips. Considering the angle of repose of  $30^\circ$  and the length of the conveyor, the maximum chip pile height is 41 ft. Assuming an angle of  $270^\circ$  is swept out by the conveyor and that the bulk density of the wood chips is  $19 \text{ lb/ft}^3$ , the chip pile storage is calculated to be approximately 4 days. The flume pond size is assumed to be  $100 \text{ ft} \times 100 \text{ ft}$ .

A building, arbitrarily sized at  $140 \text{ ft} \times 150 \text{ ft}$ , houses the disk refiners, impregnation and prehydrolysis reactors,  $\text{H}_2\text{SO}_4$  tank, blowdown tank, neutralization vessel, and the lime storage bin. The lime pit size is taken from Badger's equipment list and is located outside the prehydrolysis building and covered by a shed to protect the open lime pit. Lime is conveyed by a covered conveyor to the lime storage bin inside the building.

The fermentation area (cellulase production, xylose fermentation, and SSF) is sized by laying out each tank in the area as shown in Drawing No. BESSF-1001. The total area required is approximately  $340 \text{ ft} \times 560 \text{ ft}$ . It is assumed that 10 ft would be an adequate spacing between fermenters. The control room and laboratory buildings are also arbitrarily sized.

Detailed equipment layout is not done for the ethanol purification and solids separation area. However, based on the size of the columns and some ancillary equipment, this area is assumed to be approximately  $100 \text{ ft} \times 120 \text{ ft}$ . A  $100 \text{ ft} \times 120 \text{ ft}$  building is added to house the centrifugation equipment.

## Section 5.0

### Economic Analysis of Base Case

#### 5.1 Capital Cost

A breakdown of purchased equipment cost (1990 dollars) by process area is given in Table 5-1. The purchased cost of the boiler is not included in the utility area because this cost was for an installed package. The total purchased equipment cost (except for the boiler) is \$37.22 MM.

The installed cost of the equipment (except for the boiler) is 2.85 times the purchased equipment cost. This factor was obtained from capital cost data for existing corn-to-ethanol plants and from ICARUS (1987) information. The installed cost includes direct field costs (labor and materials for purchased equipment, equipment setting, piping, civil, steel, instrumentation, electrical, insulation, painting, buildings, and land) and indirect costs (engineering, construction, overhead, contractor's fee, contingency, and special charges). The installed cost of the equipment, plus the installed cost of the boiler package, plus a miscellaneous cost for unlisted equipment (assumed to be 2% of the fixed capital investment) gives a fixed capital investment of \$128.42 MM.

The total capital investment is the sum of the fixed capital investment, start-up cost, and working capital. Start-up cost is 5.0% of fixed capital investment (Chem Systems 1990). Working capital is calculated according to the methodology of Chem Systems (1990) and is the sum of three items:

1. Accounts receivable - one month's gross cost of production (net cost of production less by-product credit)
2. Cash - one week's gross cost of production
3. Warehouse/spares - 3.0% of fixed capital investment

Less a fourth item:

4. Accounts payable - one month's raw material cost

The total capital investment for this plant is \$141.24 MM.

#### 5.2 Economic Analysis

The total cost for denatured fuel is shown in Table 5-2 for a production rate of 58.5 MM gal/yr. Both the capital cost estimate and the production economics are done on the basis of first-quarter 1990 dollars. Chemical costs were taken from the *Chemical Marketing Reporter* (1990), wood was assumed to cost \$42/dry ton (2.1¢/lb), and makeup water was assumed to cost 10¢/1,000 gal. The solids from the boiler are disposed of at a cost of \$20/ton and a credit was taken for excess electricity produced at 4.0¢/kWh.

The economic basis for this plant is again taken from Chem Systems (1990) and is given below:

- Labor is 41 operators/technicians at \$29,800/yr
- Nine foremen at \$34,000/yr
- One supervisor at \$40,000/yr
- Direct overhead is 45% of manpower cost
- Maintenance is 3% of fixed capital investment
- General plant overhead is 65% of manpower costs and maintenance

**Table 5-1. Capital Cost Summary**

Plant Area	Base Case Equipment Cost (\$ MM)
100 Wood handling	2.51
200 Prehydrolysis	8.31
300 Xylose fermentation	
Seed fermenters	0.34
Fermenters	1.78
Remaining equipment	<u>0.04</u>
Total	2.16
400 Cellulase production	
Seed fermenters	0.09
Fermenters	0.80
Remaining equipment	<u>0.08</u>
Total	0.97
500 SSF	
Seed fermenters	1.11
Fermenters	6.17
Remaining equipment	<u>0.08</u>
Total	7.36
600 Ethanol recovery	
Rectification column	0.25
Remaining equipment	<u>1.15</u>
Total	1.40
700 Off-site tankage	1.44
800 Waste treatment	
Wastewater treatment	1.24
Remaining equipment	<u>0.15</u>
Total	1.39
900 Utilities	
Boiler feed water	1.31
Process water	0.33
Turbogenerator	6.50
Cooling water	0.92
Chilled water	0.60
Fermentation air	1.82
Auxiliary utilities	<u>0.20</u>
Total	11.68
Total Equipment Cost (except boiler)	37.22

**Table 5-1. Capital Cost Summary (Concluded)**

Total Equipment Cost (except boiler)	37.22
Times 2.85 installation factor	106.08
Boiler package	19.82
Miscellaneous	2.52
Total Fixed Capital Investment	128.42
Startup cost	6.42
Working capital	6.40
Total Capital Investment	141.24

- Insurance and taxes are 1.5% of fixed capital investment
- Annual capital charges are 20% of total capital investment. This charge is approximately equivalent to a 10% discounted cash flow rate of return with the following parameters:
  - Three years for construction with expenditures of 30% in the first year, 50% in the second year, and 20% in the third year
  - Fifteen years of operation
  - Income tax rate of 37%
  - No sales expenses
  - Capacity buildup of 60% of nameplate capacity in the first year, 80% in the second year, and 100% from the third year onward
  - Depreciation at 5 years straight line for battery limits investment and 15 years straight line for outside battery limits investment.

The variable operating cost is 60.1¢/gal and the fixed operating cost is 19.8¢/gal, giving a gross cost of production of 79.8¢/gal. When by-product credits are included, the net cost of production is 73.4¢/gal. The annual capital charge of 48.3¢/gal, added on to the net cost of production, gives a denatured fuel cost of 121.7¢/gal.

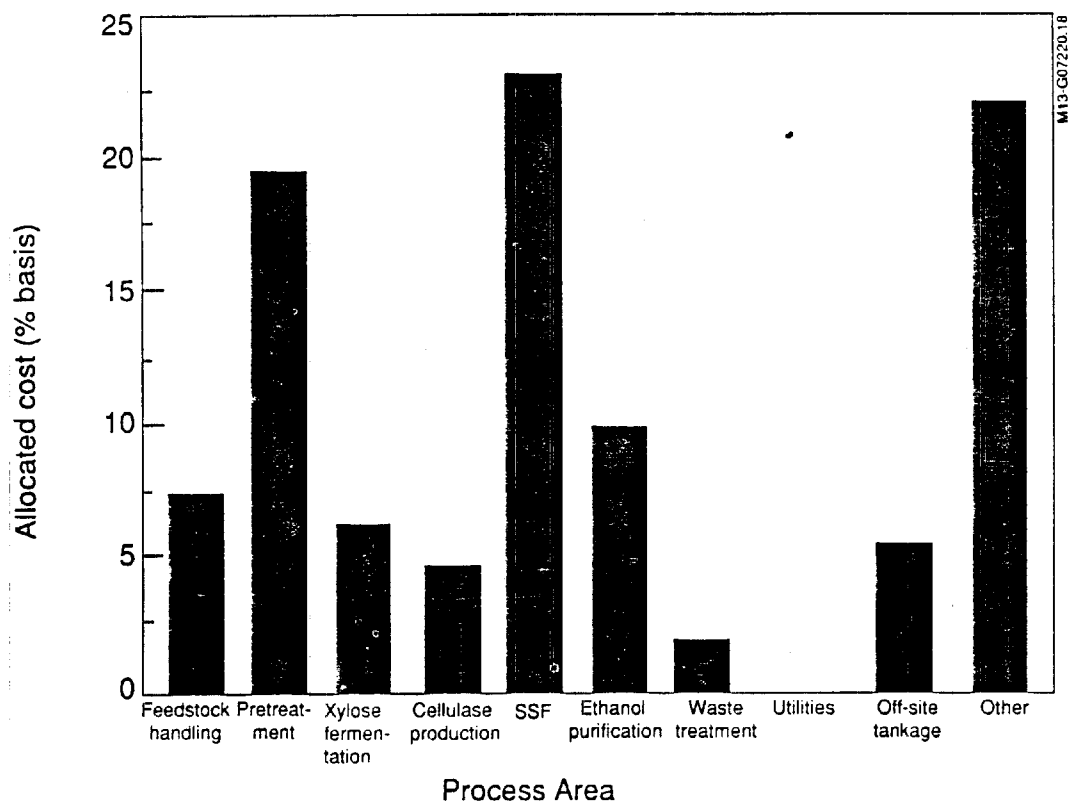
### 5.3 Allocated Cost

The allocation of total cost for utilities and feedstock to each of the plant areas using these services gives a more realistic assessment of cost for each process area. Feedstock cost is allocated based on the heating value of each feed component (i.e., cellulose, xylan, lignin, etc.) and the area of the plant that uses that component. Leftover lignin could not be allocated to any of the other areas and so was placed in the other category. The capital cost of utility services such as steam, air, chilled water, cooling water, and electricity is allocated based on demand in each plant area. The allocated cost for utilities fortuitously is zero, because the remaining cost for the turbogenerator and boiler is canceled by the credit for electricity production.

As shown in Figure 5-1, the allocated cost of SSF is the largest in the plant, followed closely by pretreatment. Therefore, these areas offer the greatest potential for cost reductions either by reducing capital and/or operating cost. Wastewater treatment, which accounts 2.2% of the total allocated cost, leaves little room for improvement.

**Table 5-2. Cost of Production Estimate for Denatured Fuel**

Capacity: 58.5 MM gal/yr denatured fuel			
Total Capital Investment: \$141.24 MM			
Capital Investment/Annual Gallon: \$2.41/gal			
	¢/unit	MM\$/yr	¢/gal
<b>Materials</b>			
Wood (lb)	2.10	26.88	45.9
Acid (lb)	3.75	0.86	1.5
Lime (lb)	2.25	0.38	0.7
Ammonia (lb)	5.00	1.76	3.0
Corn steep liquor (lb)	11.00	0.40	0.7
Nutrients (lb)	12.40	0.13	0.2
Antifoam (lb)	26.00	0.06	0.1
Glucose (lb)	53.00	1.68	2.9
Gasoline (lb)	12.81	2.21	3.8
Diesel (lb)	12.81	0.64	1.1
Makeup water (lb)	0.00	0.14	0.2
<hr/>			
Subtotal - Variable Cost		35.16	60.1
<b>Fixed Operating Cost</b>			
Labor		1.20	2.1
Foreman		0.30	0.5
Supervision		0.00	0.1
Direct overhead		0.70	1.2
Maintenance		3.90	6.6
General plant overhead		3.60	6.0
Insurance and taxes		1.90	3.3
<hr/>			
Subtotal - Fixed Operating Cost		11.60	19.8
<b>By-product credits</b>			
Electricity (kWh)	4.00	4.16	7.1
Solids Disposal (lb)	1.00	(0.4)	(0.7)
<hr/>			
Net cost of production		43.00	73.4
<hr/>			
Annual capital charge		28.20	48.3
<hr/>			
Denatured fuel cost		71.20	121.7



**Figure 5-1. Total plant cost allocated to each of the process areas**



## Section 6.0

### Optimization of Base Case Cellulase Production and SSF Performance

A series of analyses was carried out to determine the optimum cellulase loading to SSF for three different conversion levels: 72%, 80%, and 85%. SSF residence time for a given conversion level and cellulase loading combination is determined using the yield expression described in Section 3.2.4 of this report. Ethanol yields for Genencor batch I enzyme, with a mixed culture of *S. cerevisiae* and *B. clausenii* on *Populus* substrate, are used. The values used for cellulase loading and SSF fermentation time in the analysis are, for the most part, within the limits of the data. The data ranges are fermentation times of 2 to 8 days and cellulase loadings of 7 to 26 IU/g cellulose. Only the lowest cellulase loadings for the 72% conversion level are slightly outside the regressed data range.

The optimization between cellulase production and SSF performance is relatively complex. When cellulase loading is increased for a given SSF conversion level, the following major changes result:

#### Decreases in:

- SSF capacity and capital costs
- Ethanol purification capacity and costs
- Ethanol production rate
- Grand total capital investment (over the range of cellulase loadings investigated)
- Net power to sales.

#### Increases in:

- Cellulase production capacity and capital costs
- Boiler capacity and capital cost
- Turbogenerator capacity and capital cost
- Chilled water system capacity and capital cost
- Fermentation air system and capital cost
- Chemical and nutrient costs.

Figure 6-1 shows the variation of total ethanol cost as a function of cellulase loading for the three conversion levels. Each of the three curves has a minimum ethanol cost at a different cellulase loading. The estimated minimum costs are:

- 121.7¢/gal for 6.85 IU/g cellulose at 72% SSF conversion level, 7-day SSF
- 117.0¢/gal for 9.0 IU/g cellulose at 80% SSF conversion level, 7.3-day SSF
- 114.7¢/gal for 11.0 IU/g cellulose at 85% SSF conversion level, 7.3-day SSF.

These curves apply only when the balance of the base case process configuration and costs remain constant. The optimum cellulase loading for the 72% conversion level is the same as for the base case. The sensitivity analysis shows that higher cellulase loadings and higher SSF conversion levels decrease the overall cost of ethanol substantially. Conversion levels higher than 85% were not examined because it would have required changing the proportions of cellulose consumed in side reactions in the SSF system. It is interesting to note that the three optimum cellulase loadings occur at about a 7-day SSF residence time.

Thus, for the process design described for the base case and within the levels of the regressed yield data, the optimum cellulase loading is 11 IU/g of cellulose with a conversion time of 7.3 days and a yield of 85% of theoretical. However, the optimum cellulase loading is sensitive to changes in other process and economic assumptions. For example, if stainless steel fermentation vessels are required, then the

optimum cellulase loading would be at higher levels than above because SSF costs would rise faster than other costs would decrease. Any process changes that would decrease the cost of cellulase production such as reducing fermentation air and/or chilled water requirements, or reducing chemical and nutrient needs, would also shift the optimum cellulase loading levels to higher values.

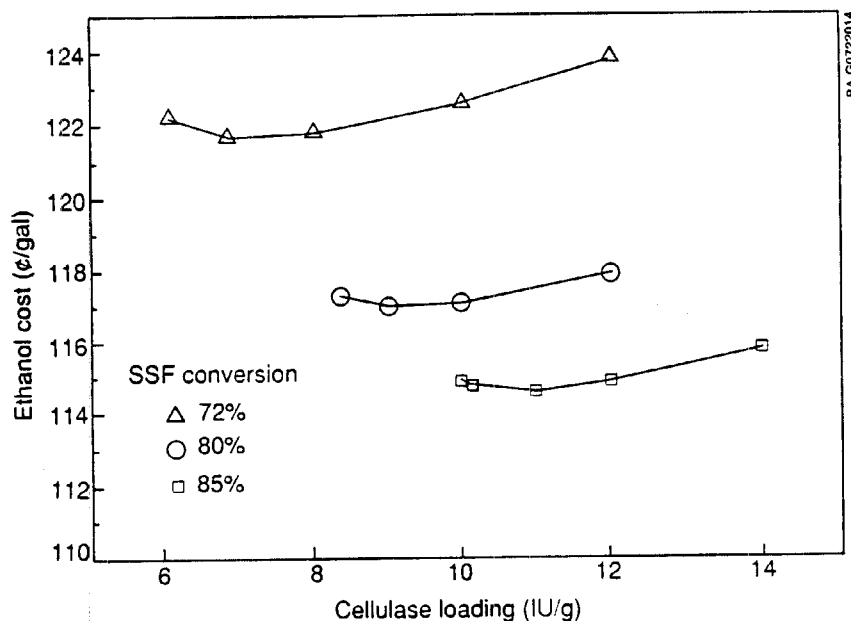


Figure 6-1. Ethanol cost as a function of SSF ethanol yield and cellulase loading

## Section 7.0

### Sensitivity Analysis

This section describes the results of sensitivity analyses carried out on the unoptimized base case design for the biomass-to-ethanol process described in Section 4.0. The purpose of these analyses is to quantify the economic effect of changes in the overall design and various economic parameters and assumptions used for the base case. For most cases, only one variable at a time is changed from the base case. However, for some cases, several variables are changed simultaneously to show the compound effect of multiple changes. The results of these analysis provide information for (1) prioritizing future R&D and analysis activities to strengthen understanding of the base case process, (2) understanding the implications of economic assumptions used in the base case, (3) prioritizing future R&D and analysis activities to establish new technologies for improving the base case economics, and (4) prioritizing future R&D and analysis activities on feedstocks. The results also show the compound effect of improvements and indicate what can ultimately be achieved in terms of production cost of ethanol.

A spreadsheet model of the SSF-based biomass-to-ethanol plant was developed to help perform some of the sensitivity analyses. The model can calculate a complete process material balance, utility balances, chemical requirement summaries, and an economic evaluation as a function of technical and economic performance parameters and assumptions. Capital costs for modified designs are ratioed from the base case costs in the economic evaluation portion of the model using appropriate capacity factors. A more complete description of this model is provided in Appendix D.

#### 7.1 Analysis Relevant to Prioritizing R&D and Analysis Activities to Strengthen Understanding of the Base Case Process

##### 7.1.1 Milling Power

A biomass feedstock must be milled prior to pretreatment. The power required for milling is a function of feedstock type, feedstock size, mill type, and final size requirement. For the base case, milling power is approximately 128 hp-h/dry ton. This is based on 1-in. wood chips fed to a disk refiner and a final size of less than 3/16 in. Power consumed in the milling operation directly reduces the power available for sale. An analysis determined the effect of different milling power levels on the amount of power available for sale and the resulting total cost of ethanol. For the base case, approximately 60 kW are required for every 1 hp-h/dry ton increase in milling power consumption. At \$0.04/kWh, this is equivalent to lost revenue of \$19,200 per year for every 1 hp-h/dry ton increase. Figure 7-1 shows how power to sales decreases and ethanol cost increases with increasing milling power requirements. The range of milling power investigated is from 20 to 200 hp-h/dry ton. The low end of the range represents power required to knife mill herbaceous crops while the high end corresponds to wood chips processed in disk refiners with a small final particle size requirement.

Feedstock milling is the single most power-intensive operation in the base case design of the biomass-to-ethanol process. More data are needed to determine the actual milling power expected for biomass preparation. This information is important for evaluating feedstocks as well as milling equipment.

##### 7.1.2 Increase Percent Solids from Neutralization

Minimizing the addition of water to the downstream tank (T-203) and thereby increasing the solids concentration would reduce the size of most of the downstream equipment in the plant. Twelve percent solids is selected because it is believed that this concentration could be pumped with current technology.

However, if the maximum solids concentration is only 8%, the ethanol concentration out of SSF would be 2.9% compared to 4.35% for the base case. The ethanol cost would rise to 136.9¢/gal, an increase of 15.2¢/gal compared to the base case. On the other hand, if the solids concentration is 15%, the ethanol concentration from SSF would be 5.4% and the ethanol cost would drop to 115.5¢/gal, a decrease of 6.2¢/gal compared to the base case. If dilution water is added to SSF due to tolerance limits, at least part of the advantage of higher solids concentration from neutralization would be eliminated. These effects on the different solids concentrations are shown in Figure 7-2.

From this analysis, it is clear that minimizing water addition to the flash tank while maximizing solids content is desirable as long as the material can be pumped. Experimental work is required to determine the maximum solids concentration that can be pumped.

### 7.1.3 Nutrients for SSF and Xylose Fermentation

In the base case, all fermentations, except cellulase production, are assumed to require no additional nutrients above what is contained in the recycled process water. If nutrients are required, they could add significantly to the cost of ethanol. Appendix A-6 presents an example of potential nutrients required with an associated cost of 35.2¢/gal of ethanol product (14.2¢/gal for SSF nutrients and 21.1¢/gal for xylose fermentation nutrients). This would represent a 28% increase in ethanol cost to 156.9¢/gal.

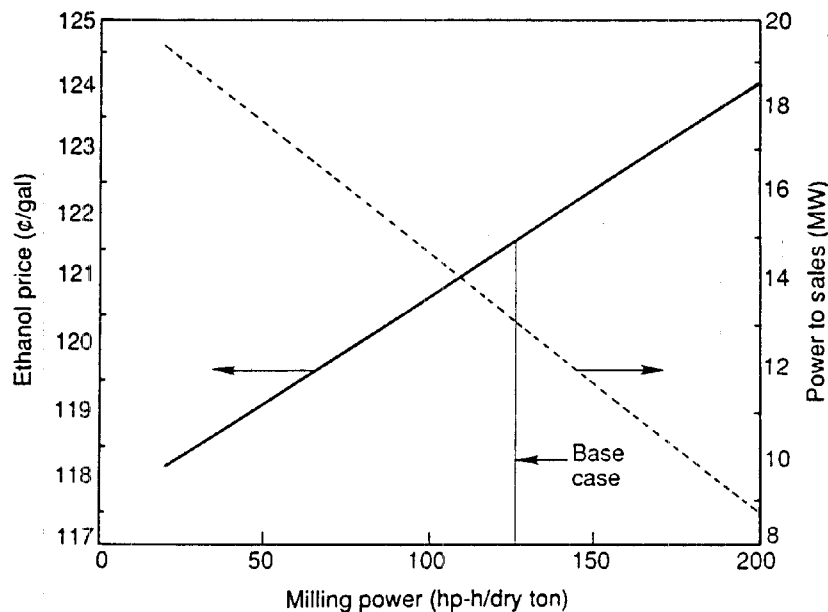
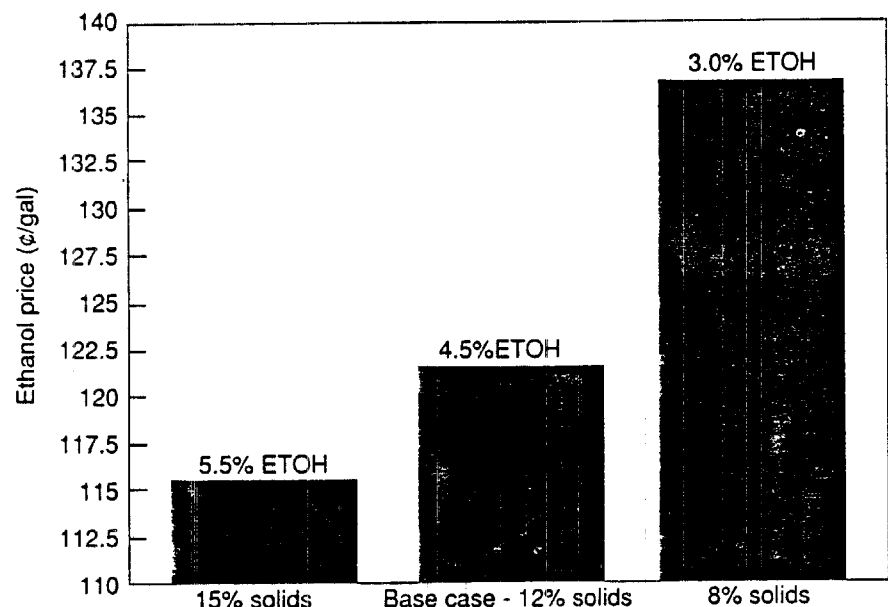


Figure 7-1. Ethanol price and power export to sales as a function of milling power



**Figure 7-2. Ethanol price as a function of solids concentration allowing ethanol concentration to vary**

### 7.1.4 Fermentation Stirring Power

Mixing power requirements in xylose fermentation and SSF are currently unknown. A level of 0.1 hp/1,000 gal is assumed for the base case. The current fermentation design scheme in both xylose fermentation and SSF calls for a continuous flow through large fermenters arranged in series. It may be that the flow of fluid through the vessels combined with CO<sub>2</sub> evolution during fermentation may result in sufficient fluid mixing. However, if additional stirring is required, the power input necessary could have a large impact on the overall plant power balance.

Mixing power for xylose fermentation and SSF, both main fermenters and seed fermenters, are varied between 0.1 and 2.0 hp/1,000 gal. Figure 7-3 shows how ethanol cost and power to sales vary as a function of mixing power requirements. At a mixing power input of 0.7 hp/1,000 gal, there is no excess power available for sale; at levels above this, additional power would have to be purchased or generated on site. The ethanol cost curve shown in Figure 7-3 assumes that power can be sold or bought for \$0.04/kWh. In reality, the cost to purchase power is probably higher than the price a utility would pay for power so that the slope of the ethanol cost curve might actually increase as mixing power rises above the 0.7 hp/1,000 gal threshold.

Identifying the minimum stirring power requirements for xylose fermentation and SSF is very important for the optimization of the biomass-to-ethanol process. In the base case design, the total fermenter volume in xylose fermentation and SSF is about 27 MM gal. An increase in stirring power of 0.1 hp/1,000 gal results in an increase in internal power consumption of 2,700 hp. With power at \$0.04/kWh, this is equivalent to lost revenue of about \$644,000 per year for every 0.1 hp/1,000 gal increase.

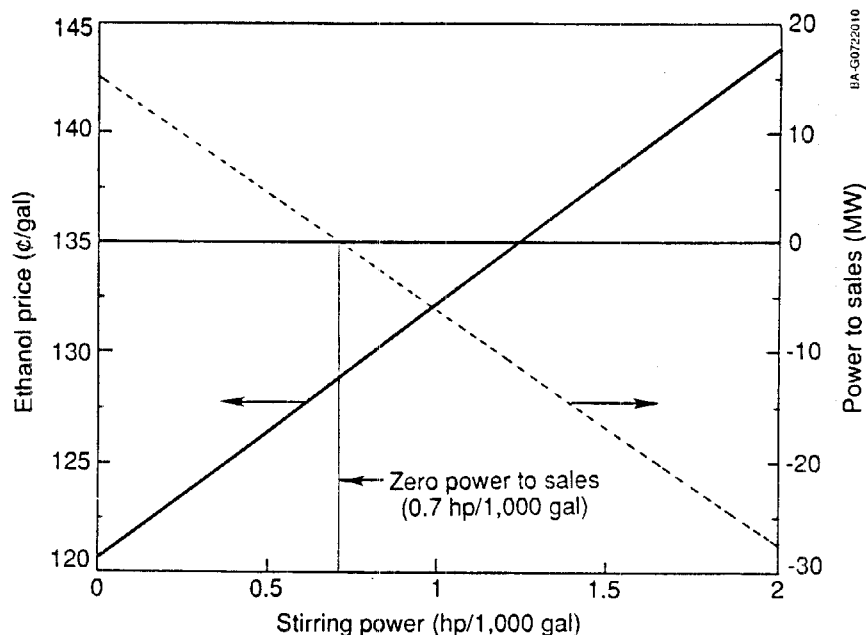


Figure 7-3. Ethanol price and power export to sales as a function of fermenter stirring power

### 7.1.5 Process Water Recycle

A partial process water recycle is incorporated into the base case design where part of the process water from the lignin centrifuge is recycled back to the front end of the plant as dilution water. This decreases the size of the wastewater treatment system and makeup water requirements. It also has the effect of returning unconverted xylose and nutrients back to the fermentations. Furthermore, water from the centrifuge is at a temperature of 100°C and can be presumed to be sterile. Thus, use of this water in the process will minimize introduction of outside microbial contamination. The major disadvantage is that certain components of the feed and reaction by-products, which would normally stay in the liquid phase, will accumulate to a higher concentration until what is entering or being produced in the process is offset by losses in streams to wastewater treatment and the boiler. Soluble solids and ash in the biomass feed and reaction by-products such as glycerol fall into this category.

Sensitivity analysis indicates that if no water is recycled, and only fresh water is used, the cost of ethanol would increase by 4.9¢/gal to 126.6¢/gal compared to the base case. The increase is due to a 3.4% drop in ethanol production from xylose, and increased capital costs in wastewater treatment and steam and power production. Additional nutrient costs are not included. Increased revenues from electricity sales from burning the extra xylose helps to offset to some extent some of the higher costs. On the other hand, when all dilution water is provided by recycle water, the cost of ethanol drops by 3.7¢/gal to 118.0¢/gal compared to the base case. The ethanol production rate is 2.9% higher than that of the base case because of the availability of extra xylose. Capital costs for wastewater treatment and steam and power production, as well as electricity revenues, are all lower. These effects on ethanol cost of the various recycle scenarios are shown in Figure 7-4. Experimental work is required to determine the amount of recycle that can be employed without producing significant negative impact on the process, particularly fermentations, and to determine if there is nutrient value to the recycle water.

### 7.1.6 Materials of Construction

All fermentation vessels are assumed to be of carbon steel in the base case cost estimate. Experience in the corn-to-ethanol industry has been primarily with stainless steel fermentation vessels. If stainless steel equipment is required in xylose fermentation, cellulase production, and SSF, the fixed capital investment would increase by as much as \$28.2 MM with a resulting increase in ethanol cost of 13.1¢/gal to 134.8¢/gal compared to the base case.

This cost for ethanol is determined by using an installation factor of 5.4 for all the equipment in xylose fermentation, cellulase production, and SSF. This factor is arrived at by considering the increase in purchase price for stainless steel equipment and the increased piping cost, which increases the direct installed cost. Because contingencies and fees are determined from the direct installed cost, they also increased for the stainless steel case. It is assumed that the cost for other factors—concrete, instruments, electrical, insulation, paint, labor, indirects (freight, taxes, insurance, construction overhead, engineering expenses), and auxiliary facilities—would be the same as the carbon steel case. The breakdown is shown in Table 7-1, which contains information from Ulrich (1984). The installation factor for carbon steel in the table is 3.41, and since we used 2.85, a ratio converts the 6.66 factor of this table into the approximate 5.4 ( $3.41/2.85 \times 6.54$ ) factor we used.

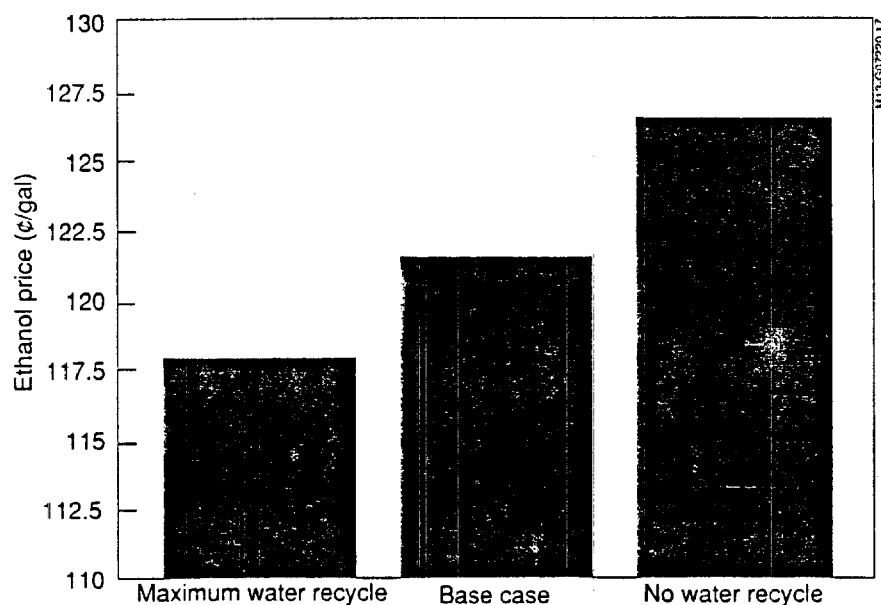


Figure 7-4. Ethanol price as a function of water recycle rate

This analysis showed the desirability of using low-cost materials of construction. A more detailed analysis is required to determine whether carbon steel can in fact be used in SSF and other areas. Such an analysis should also look into the possibility of using other low-cost materials such as plastics.

### 7.1.7 Operational Reliability

Overall plant reliability is usually measured as percent of annual onstream time. For the base case, the onstream time is 8,000 h/yr or 91.3%. Figure 7-5 shows ethanol cost as a function of onstream time. At 85%, 7446 h/yr, ethanol cost is 127.7¢/gal, while at 100%, 8760 h/yr, ethanol cost drops to 115.8¢/gal, relative to the base case cost of 121.7¢/gal. This sensitivity illustrates the importance of designing reliability into the plant and striving toward 100% onstream time.

**Table 7-1. Purchased Equipment Installation Factor**

	Carbon Steel	Stainless Steel
Purchased equipment (PE)	100.0	335.0
Piping	12.8	42.9
Concrete	8.6	8.6
Steel	-	-
Instruments	3.8	3.8
Electrical	0.6	0.6
Insulation	-	-
Paint	1.2	1.2
Total material	127.0	392.1
Labor (PE × 0.267)	26.7	26.7
Direct installed cost (DIC)	153.7	418.8
Indirects (DIC × 0.26)	40.0	40.0
Total module cost (TMC)	193.6	458.8
Contingency (TMC × 0.15)	29.0	68.8
Fee (TMC × 0.03)	5.8	13.8
Auxiliary facilities (PE × 1.13)	113.0	113.0
Grand Total	341.4	654.4

Vendor testing of large-scale equipment with materials actually used in the process, together with the operation of a process development unit (PDU) of key unit operations, is necessary to achieve a realistic understanding of the actual reliability of the process. Eventually, operation of an experimental demonstration unit (EDU) is required to better establish reliability.

### 7.2 Analysis of Economic Assumptions Used in the Base Case Analysis

An installation factor of 2.85 is applied to all purchased equipment costs, except the boiler, to arrive at the total fixed investment cost of the plant. Because the boiler price used is an installed cost, it did not require this factor. Figure 7-6 shows ethanol cost as a function of the value of the installation factor. At a factor of 2.5, ethanol cost is about 115.3¢/gal, while at a factor of 3.5, ethanol cost is 133.6¢/gal. Overall installation factors for processing plants are a function of many variables including size, materials of construction, operating conditions, type of process control, and environmental requirements. Installation factors for petrochemical plants and refineries typically are in the range of 2.5 to 5.0. An installation factor in the low end of this range for the biomass-to-ethanol process is reasonable because the process is simple, takes place at mild operating conditions, and should have minimal environmental constraints.



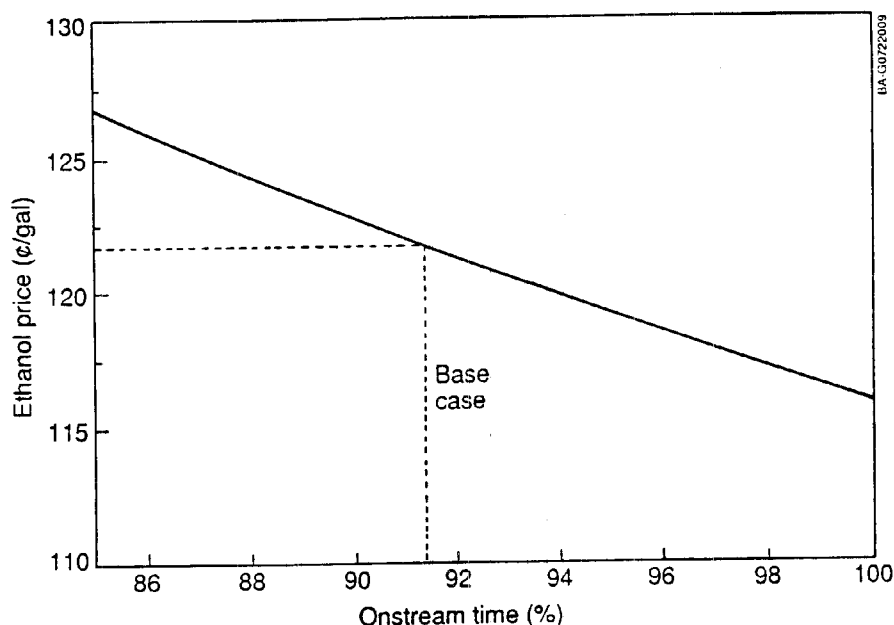


Figure 7-5. Ethanol cost as a function of onstream time

### 7.3 Analysis Relevant to Prioritizing R&D Activities to Establish New Technologies that ~~Will Improve the Recycle Economics~~

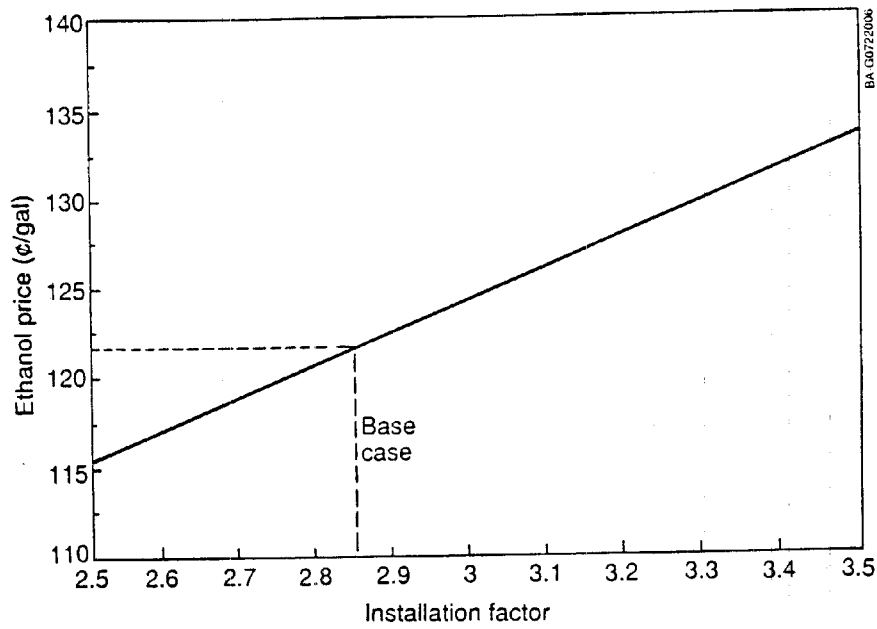
#### 7.3.1 Increase Xylan to Xylose Conversion in Pretreatment

Using the spreadsheet model, it was determined that if xylan to xylose conversion in pretreatment is increased from 80% to 90%, the cost of ethanol will drop by 3.5¢/gal from 121.7¢/gal to 118.2¢/gal (see Figure 7-7). The remaining xylan is assumed converted to furfural, which ultimately ends up in the flash vapor stream.

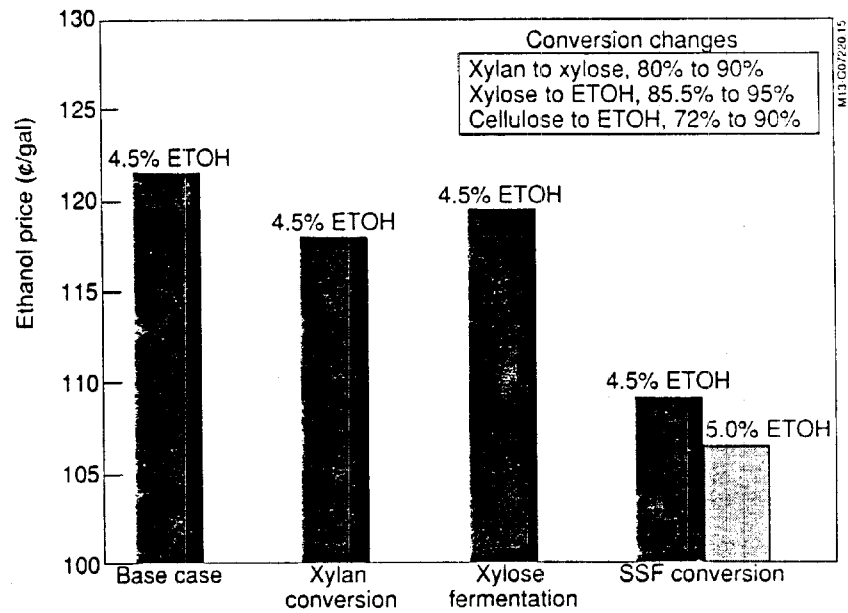
It is important to remember for this analysis, that the base case includes partial water recycle. Also in the base case, it is assumed that 80% of the xylan remaining after pretreatment is converted to xylose by xylanases present in SSF. As a result, some of the xylan unconverted in pretreatment is ultimately converted to ethanol because xylose made in SSF from the xylan is recycled back to the front end of the plant and eventually reaches the xylose conversion unit. Thus, improving xylan conversion in pretreatment is most important when there is no process water recycle and least important at maximum process water recycle. Thus, if experimental work indicates that maximum process water recycle is feasible, the importance of increasing xylan to xylose yield may be minimal. On the other hand, if no recycle is possible, the effect of increased xylan to xylose yield would be greater than the 3.5¢/gal indicated above.

#### 7.3.2 Increase Xylose to Ethanol Conversion in Xylose Fermentation

In the base case, it was assumed that only 90% of the sugars (xylose and glucose) is available for conversion to ethanol in xylose fermentation because all the sugars do not fully diffuse from the wood particles during the course of the fermentation. Of the available sugars, 95% are converted to ethanol and with 90% assumed to diffuse out of the wood, this gives an overall yield of 85.5% for the base case.



**Figure 7-6. Ethanol cost as a function of the installation factor**



**Figure 7-7. Ethanol price for changes in xylan to xylose, xylose to ethanol, and cellulose to ethanol conversion levels**

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Using the spreadsheet model, it was determined that if 100% of the sugars are available and with 95% conversion to ethanol, the ethanol cost would drop by about 2.4¢/gal to 119.4¢/gal (see Figure 7-7). The drop is due to a 2.2% increase in ethanol production from the extra available xylose. Capital cost decreases caused by smaller boiler and turbogenerator are offset by reduced revenues from electricity sales.

### **7.3.3 Increase Cellulose to Ethanol Conversion in SSF**

Using the spreadsheet model, it was determined that increasing conversion of cellulose to ethanol in SSF from the base case value of 72% to 90%, while keeping cellulase loading and residence time the same as in the base case, would result in a 15.0¢/gal drop in ethanol cost to 106.7¢/gal. However, at this increased conversion and with no dilution water added to SSF, the concentration of ethanol in the stream from SSF is 5.04 wt %, which is greater than the ethanol tolerance limit of 4.5 wt % that is used in the base case. When dilution water is added to keep the ethanol concentration at the base case tolerance limit, the cost of ethanol would be 109.4¢/gal, a decrease of 12.3¢/gal from the base case (see Figure 7-7). The cost increase that results from addition of dilution water is due to a lower ethanol production rate as well as increased capital costs for SSF, ethanol recovery, wastewater treatment, and lower revenues from power sales. Lower ethanol production results from higher substrate losses to the boiler and wastewater treatment because of the additional water required for dilution while maintaining a constant process water recycle rate.

### **7.3.4 Reduce Fermentation Times**

Using the spreadsheet model, the effect of decreased fermentation times for xylose fermentation, SSF, and cellulase production on the price of ethanol was investigated. Figure 7-8 shows how ethanol cost decreases from the base case value as fermentation time decreases for xylose fermentation, SSF, and cellulase production. All other variables are the same as the base case.

Decreasing xylose fermentation time from the base case value of 2 days to 1 day will decrease ethanol cost by 1.3¢/gal to 120.4¢/gal. This decrease is due to lower capital costs and stirring power requirements in the xylose fermentation area.

SSF fermentation time for the base case is 7 days. Ethanol costs drops by 2.7, 4.7, and 6.7¢/gal for SSF fermentation times of 5, 3.5, and 2 days, respectively. As with xylose fermentation, decreasing ethanol costs are due to lower capital investment and stirring power requirements. Cost savings are substantial because SSF fermentation accounts for about 19.8% of the purchased equipment cost (excluding boiler) in the base case plant, compared with 5.8% for xylose fermentation and 2.6% for cellulase production.

Cellulase production is a batch process with a total cycle time of 6 days in the base case design. If cycle time is reduced to 4 days and 2 days, ethanol cost drops by 0.6 and 1.3¢/gal, respectively. The cost reduction is relatively small because only a small slipstream of the neutralized hydrolyzate is used as the feed to cellulase production.

### **7.3.5 Eliminate Xylose and SSF Seed Fermentations**

It is assumed that all the seed fermentations, xylose, SSF, and cellulase, are aerobic. This is an important assumption because the fermentation air compressor and the required chilled-water capacity account for about 4.5% of the total fixed capital investment and more than 20% of the internal power demand. Growing the *E. coli* inoculum for the xylose fermentation consumes more than 4% of the hydrolyzed biomass as well as a large quantity of fermentation air. Dilution water is also added to the xylose seed fermenters to reduce the glucose plus xylose concentration in the feed to the seed fermenters

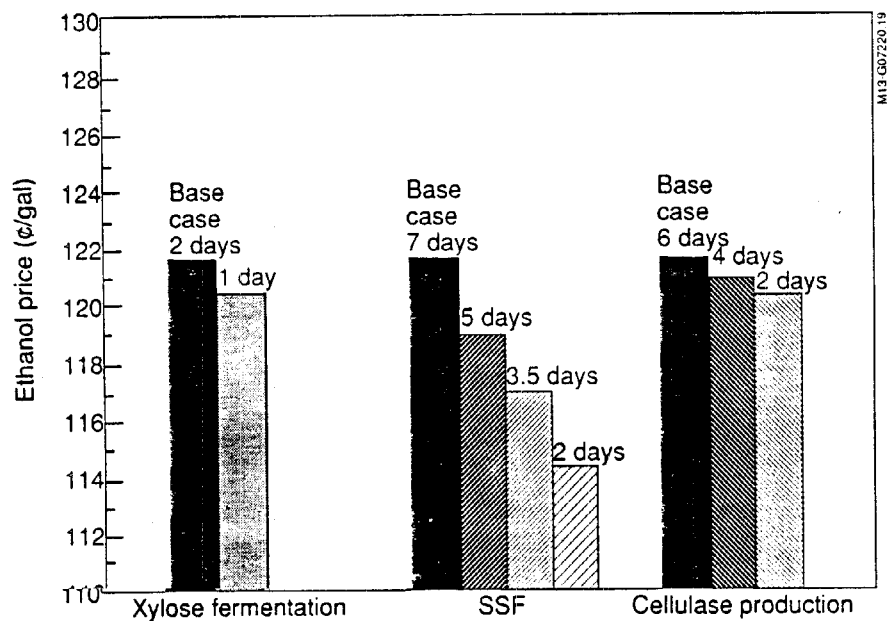


Figure 7-8. Ethanol price for different fermentation rates in xylose fermentation, SSF, and cellulase production

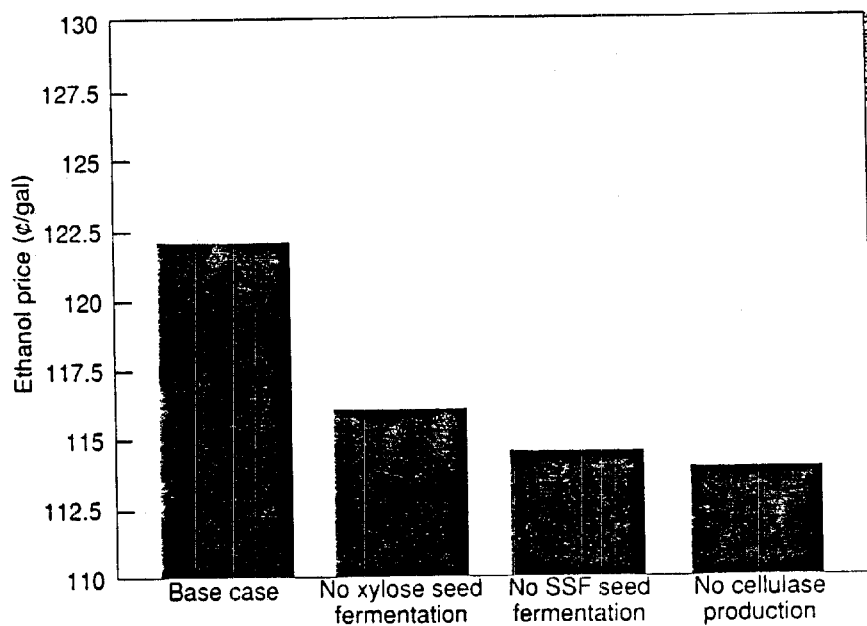


Figure 7-9. Ethanol price after elimination of xylose seed fermentation, SSF seed fermentation, and cellulase production

to 2 wt % and to get to an inoculum level of 10%. The elimination of the seed fermenters reduces the total volumes required for xylose fermentation and SSF and the ethanol production rate is increased by more than 2.5%, resulting in ethanol cost falling 5.5¢/gal to 116.2¢/gal (see Figure 7-9). The intent of this sensitivity analysis is to show the maximum incentive for modifying the overall process such that separate xylose seed fermentation is no longer required. For example, an *E. coli* recycle system will result in additional capital and operating cost. If the new system added more than 5.5¢/gal to the cost of ethanol, then the process modification would not be made.

Eliminating SSF seed fermentation could potentially reduce the cost of ethanol by about 7.0¢/gal to 114.7¢/gal (see Figure 7-9). Ethanol production increases by just over 2%, and capital cost drops by almost 5%. The drop in capital is because elimination of the seed fermentation equipment and reduction in size of the fermentation air system. In addition, revenues from electricity sales are increased because of lower internal power demand, caused by elimination of agitators and reduction in fermentation air capacity. As with the analysis for eliminating xylose seed fermentation, the intent of this analysis is to show the maximum incentive for modifying the process to eliminate SSF seed fermentation.

### **7.3.6 Eliminate Cellulase Production**

This sensitivity analysis is to provide information for a make or buy decision concerning cellulase enzyme. The information is also relevant to determining if cellulase recycling is important. Cellulase production requires about 2% of the hydrolyzate. Eliminating cellulase production will reduce ethanol cost by about 7.5¢/gal to 114.2¢/gal (see Figure 7-9). The cost decrease is due to an increase in ethanol production of 2.3% and a reduction in fixed capital costs of 3.7%. An enzyme producer would have to provide cellulase at less than 7.5¢/gal of ethanol product for it to make economic sense to purchase rather than make enzyme.

## **7.4 Analysis Relevant to Prioritizing Future R&D and Analysis Activities Related to Feedstock Production**

### **7.4.1 Plant Size**

The optimum size for a biomass-to-ethanol plant is determined by the tradeoff of the economies of scale for the central processing facility versus transportation costs for the biomass feedstock. For this evaluation, the base case plant with a biomass processing capacity of 1,920 dry tons/day is compared to a larger plant with a capacity of 10,000 dry tons/day. The capacity difference is a factor of about 5.2. For both cases, biomass feedstock transportation costs are estimated. Capital and operating costs for the large plant are scaled from the base case. The total cost of ethanol is then calculated to determine where the optimum plant size may be relative to the two sizes evaluated.

As plant size increases, biomass must be collected from larger areas. If it is assumed that the collection area is circular with the plant at the center, the amount of biomass collected can be determined by the following formula:

$$\text{Biomass} = \pi \times r^2 \times \text{Productivity} \times 640$$

where

- biomass = total biomass collected in dry tons/yr
- r = collection radius in miles
- productivity = biomass produced in dry tons/acre-yr
- 640 = the conversion factor for acres to square miles.

If a mean annual productivity of 5 dry tons/acre-yr is achieved for the energy crop supplying biomass, the collection radii for the base case and large plant are about 8.4 mi and 19.1 mi, respectively. DOE research on short-rotation woody energy crops and herbaceous energy crops has already reached productivities at this level. The goal for the Short Rotation Woody Crop Program (SRWC) is 10 dry tons/acre-yr. As productivity increases, the collection radius becomes smaller, and the difference between the collection radii of the base case plant and large plant also gets smaller.

Efficient long distance transportation is important when using energy crops for ethanol production. The most important variable associated with hauling costs is hauling distance. Distance takes on extra importance since the material is of low value and low density. Although hauling costs increase with distance, the hauling cost rate (expressed in \$/ton-mi) decreases. At shorter distances, a higher percentage of the time is spent in the loading and unloading tasks. Figure 7-10 shows hauling rates for wood chips as a function of one-way travel distance. For this analysis, it is assumed that the average hauling distance is equal to the collection radius. Using large chip vans, hauling cost for the base case collection radius will be about \$0.45/dry ton-mi or \$3.78/dry ton. For the large plant, the hauling costs would be about \$0.30/dry ton-mi or \$5.73/dry ton. The hauling cost will therefore be only about \$1.95/ton higher for the large plant than the base case plant.

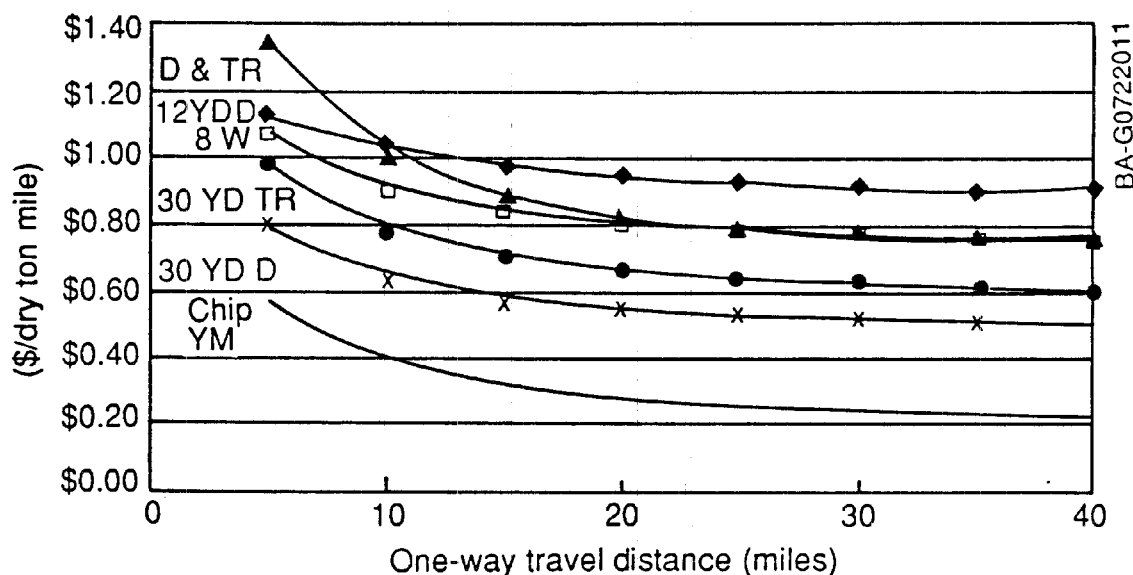
In scaling up from the base case to the larger plant, it is assumed that no economies of scale existed for the capital equipment in the following areas: wood handling, prehydrolysis, xylose fermentation, cellulase production, SSF, and off-site tankage. This is a conservative assumption, because it is expected that there will be some economies of scale as we develop an understanding of how best to scale up this equipment. The only plant areas that are assumed to have economies of scale are ethanol recovery, the environmental systems, and all equipment in the utility area. The equipment and processes in these areas are known to exhibit economies of scale. They come from existing technology in the chemical, waste treatment, and public utility industries.

In terms of operating costs, it is assumed that operating labor and supervision for the large plant would be 2.5 times that for the base case. Other fixed operating costs, such as maintenance, overhead, insurance, and taxes are estimated as a proportion of capital investment and/or labor. The same factors are used in each case. Feedstock cost is assumed to be \$42/dry ton for the base case plant and \$44/dry ton for the large plant. The extra \$2/dry ton takes into account the increased hauling distance as described previously.

The information in Table 7-2 summarizes the capital and operating costs of the two plants.

**Table 7-2. Cost Summary of Base Case and Larger Capacity Biomass-to-Ethanol Plants**

	Base Case	Large Plant
Feedstock rate (dry ton/days)	1,920	10,000
Feedstock rate (dry ton/yr)	700,800	3,650,000
Ethanol product rate (MM gal/yr)	58.5	305
Total capital investment (MM \$)	141	606
TCI/annual gallon (\$/gal/yr)	2.41	1.99
Gross COP (MM \$/yr)	46.7	233.6
Gross COP (¢/gal)	79.8	76.6
By-product credits (MM \$/yr)	3.7	19.5
By-product credits (¢/gal)	6.4	6.4
Annual cap. charge (MM \$/yr)	28.2	121.3
Annual cap. charge (¢/gal)	48.3	39.8
Total ethanol cost (¢/gal)	121.7	110.0



**Figure 7-10. Hauling cost for wood chips as a function of one-way travel distance**

From this analysis, it is evident that increased transportation costs associated with a plant size of up to 10,000 dry ton/day are more than offset by the processing plant economies of scale, resulting in significantly lower total ethanol product cost for the large plant size.

Additional analyses should be conducted in the future to identify optimum plant size as a function of biomass productivity, hauling costs, and plant economies of scale. For now it is important to recognize that transportation costs will not necessarily limit biomass-to-ethanol plants to small scales.

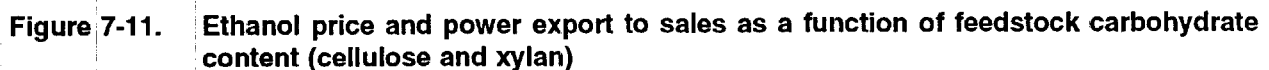
#### **7.4.2 Feedstock Composition**

The composition of the feedstock is a major variable in the analysis of the biomass-to-ethanol process. The composition used in the base case, 70.2% total carbohydrate (dry basis), is consistent with the composition of currently available hardwoods. Two sensitivity analyses are conducted at higher carbohydrate concentrations. Water in the feed is assumed to be constant at 50%, but cellulose and xylose are increased by 10%, to 77.22% total carbohydrate (dry basis), in the first case; and 20%, to 84.24% total carbohydrate, in the second case. The ratio of cellulose to xylose is maintained constant. Soluble solids, ash, and lignin concentrations are decreased but kept in the same relative proportions. Figure 7-11 shows how ethanol cost and exported power vary as a function of carbohydrate content. Exported power is important because lignin in the feed is the main fuel source for providing the internal heat and power requirements of the plant. As carbohydrate content increases, less lignin is available for fuel. At a carbohydrate content of about 82.5%, there is enough fuel in the lignin waste and by-product streams to satisfy the internal energy demand. If one goal of the process design is to require no power or fuel from outside sources, then the incentive to find higher carbohydrate feedstocks is limited by the overall energy balance. In addition, at higher carbohydrate conversion levels, it is not necessary to have such high carbohydrate concentrations in the feed to maintain the overall energy balance. It is also important to recognize that as the carbohydrate in the feed increases, ethanol concentration in the product also increases to levels above the current limit of ethanol tolerance of 4.5%. If dilution water is required to maintain

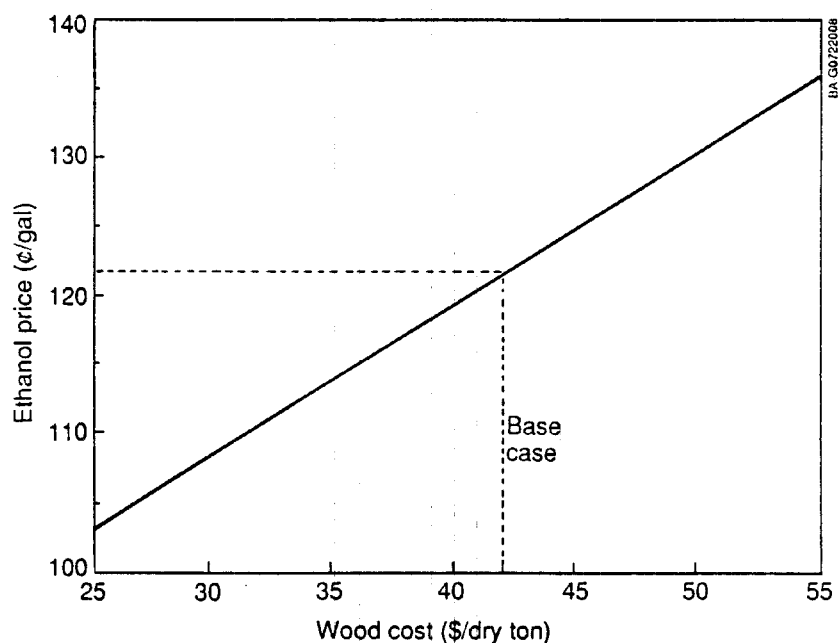
### 7.4.3 Feedstock Cost

## 7.5 Compound Effects of Process Improvements

Assuming that the process goals can be met, the spreadsheet model gives an ethanol price of 66.5¢/gal. All other assumptions for this case are the same as the base case design.







**Figure 7-12. Ethanol price as a function of wood cost**

The parameters and assumptions changed for this analysis along with base case values are shown in Table 7-3. These improvements are grouped in three categories: (1) technology improvements such as yields and productivities, (2) feedstock production improvements, and (3) more optimistic economic and operating assumptions.

**Table 7-3. Process Parameters and Assumptions for Base Case and Future Goals**

	Base Case	Goal
<b>Technology Improvements</b>		
Cellulose to ethanol yield (%)	72	90
Xylose to ethanol yield (%)	85.5	95
Xylan to xylose yield (%)	80	90
SSF fermentation time (d)	7	3
Xylose fermentation time (d)	2	1
Ethanol concentration in SSF (%)	4.35	6.6
Ethanol recovery (%)	97.0	99.8
Cellulase loading (IU/g)	7	3
SSF and xylose seed fermentations	YES	NO
<b>Feedstock Production Improvements</b>		
Feedstock cost (\$/dry ton)	42.0	34.0
Feedstock carbohydrate content (%)	70.2	77.2
<b>Economic and Operating Assumptions</b>		
Onstream time (%)	91.3	98.0
Electricity selling price (¢/kWh)	4.0	6.0
Water recycle	Partial	Maximum

## Section 8.0

### Conclusions

The economic analysis of the base case biomass-to-ethanol plant gave a cost for ethanol of 121.7¢/gal in 1990 dollars. This number is based on the cost of the equipment shown on the process flow diagrams and in the equipment list of Sections 4.3 and 4.5, respectively, with a capital recovery factor of 20%. Operating costs were generated from the material balance shown on the process flow diagrams and current costs for chemicals and utilities. Then major process variables were varied to determine their effect on the cost of ethanol. This identifies research areas that will have the greatest impact on the cost of ethanol.

Based on the sensitivity analysis presented in Section 7.0, Table 8-1 shows the effect of the various process variables on the cost of ethanol. From this table, it is easy to see which variables have the most significant positive or negative impact on ethanol cost. The table reflects only the change in ethanol cost when the given variable is varied from its base case value, while all other variables remain at their base case values. In general, the greater the impact on the process, the higher the priority should be for research on that item. However, to meet the ambitious price goal set for the biomass-to-ethanol program, any variable that decreases the cost by 1.0% or greater should be studied.

In feedstock handling, the base case value of 128 hp-h/dry ton is probably the minimum for reducing wood chips to 3/16-in. particles. If smaller particles are required, the increased power requirement (200 hp-h/dry ton) could substantially increase ethanol cost. However, herbaceous and agricultural residues have lower power requirements (approximately 20 hp-h/dry ton, Himmel et al. 1986) that could reduce ethanol cost. Different mills may be required for the different feedstocks, which would change capital cost, but this was not considered in the above analysis.

In the area of pretreatment, increasing the xylan to xylose yield to 90% will significantly decrease the cost of ethanol. However, the impact of increasing the yield will be lessened by increasing the amount of recycle water, because xylose produced from the unhydrolyzed xylan during SSF is returned to xylose fermentation. The current 80% yield has been obtained in the laboratory (Torget et al. 1988); therefore, research in this area needs to focus on increasing the yield. It will also be necessary to verify the current and future yields on a larger scale apparatus.

The solids concentration in the neutralization vessel is assumed to be 12% for 3/16-in. particles. As the sensitivity analysis shows, a change in solids concentration has a large impact on ethanol cost. The solids concentration is set at a level that results in a slurry that can be handled by downstream processes. This is dependent on particle size; smaller particles are easier to pump and mix than larger particles. Thus, there is a tradeoff between power required for size reduction and handling characteristics of the slurry. It may be necessary to decrease the particle size (i.e., increase milling power) to increase the solids concentration, thus improving the handling characteristics of the slurry.

**Table 8-1. Effect of Various Process Variables on Ethanol Cost**

Variable	Change in Variable	% Change in Ethanol Cost
<b>Feedstock Handling</b>		
Milling power	128 to 200 hp-h/dry ton	+2.3
	128 to 20 hp-h/dry ton	-3.2
<b>Pretreatment</b>		
Xylan to xylose conversion	80% to 90%	-2.9
<b>Neutralization</b>		
Solids concentration	12% to 15%	-5.1
	12% to 8%	+12.5
<b>Xylose Fermentation</b>		
Xylose to ethanol conversion	85.5% to 95%	-2.9
Fermentation time	2 days to 1 day	-1.1
Nutrients	none to Appendix C-6 data	+17.3
Seed fermentation	to none required	-4.5
Stirring power	0.1 to 2.0 hp/1,000 gal	+3.8
Materials of construction	CS to SS	+2.4
Cellulase Production	to none required	-6.2
Fermentation time	6 days to 4 days	-0.5
	6 days to 2 days	-1.1
<b>SSF</b>		
Fermentation time	7 days to 5 days	-2.2
	7 days to 3.5 days	-3.9
	7 days to 2 days	-5.5
Nutrients	none to Appendix C-6 data	+11.5
Ethanol yield	72% to 90%	-12.3
Ethanol tolerance (at 90% yield)	unlimited to 4.5%	+10.1
Materials of construction	CS to SS	+8.4
Stirring power	0.1 to 2.0 hp/1,000 gal	+13.3
Seed fermentation	to none required	-5.8

**Table 8-1. Effect of Various Process Variables on Ethanol Costs (Concluded)**

Variable	Change in Variable	% Change in Ethanol Cost
Other Factors		
Water recycle	Base value to maximum	-3.0
	Base value to none	+4.0
Plant size	1,920 to 10,000 tons/day	-9.6
Onstream time	91.3% to 85%	+4.9
	91.3% to 100%	-4.8
Wood cost	\$42 to \$55/dry ton	+11.8
	\$42 to \$25/dry ton	-15.4
Wood composition	10% increase in carbohydrates	-7.5
	20% increase in carbohydrates	-13.9
SSF/cellulase optimization	6.85 IU/g (72% yield) to	
	11.0 IU/g (85% yield)	-5.8

In xylose conversion, the single most important factor is nutrient cost. Using the media specified in Appendix C-6 will increase the cost of ethanol by 17.3%. Defining an alternative and/or minimal media, also considering the nutritive value of the recycle stream, is extremely important. The elimination of seed fermentations and increasing the xylose to ethanol yield both significantly decrease the cost of ethanol. The cost of ethanol will increase if the current assumptions for stirring power (0.1 hp/1,000 gal) and carbon steel for the fermentation tanks are not met. The least important item is decreasing the fermentation rate from 2 days to 1 day, because xylose fermentation tankage is a small part of the overall cost of the plant.

A significant cost reduction is possible if cellulase production could be eliminated or reduced by either cellulase recycle and/or increasing the activity of the enzyme. This is possible because the cellulase used for cellulase production could then be used for ethanol production. Reduction of fermentation times has little effect because again fermentation tankage is a small part of the overall cost of the plant.

In SSF, there are four variables that have a large influence (greater than 10%) on the cost of ethanol. Both nutrient addition and stirring power have a large effect because of the large SSF fermentation volume that is treated or mixed. An increase in ethanol yield can significantly reduce ethanol cost but only if ethanol tolerance is increased. Otherwise, additional volume is required to dilute the ethanol to 4.5%, offsetting the increased revenue from the extra ethanol. All the remaining factors are also important, but to a lesser extent. A 8.4% increase in ethanol cost would result if carbon steel or some other low-cost material could not be used for the fermenters. Combined with the material of construction for the xylose fermenter, this factor could account for a 10.6% increase in the cost of ethanol. Again, because of the large fermentation volume in SSF, decreasing the fermentation time or eliminating or reducing seed fermentations could significantly reduce the cost of ethanol.

Other factors having a significant impact on ethanol cost are wood composition and cost. Increasing wood cost to \$55/dry ton increases ethanol cost by 11.8%, while a wood cost of \$25/dry ton reduces cost by 15.4%. Increasing the carbohydrate content of the wood increases ethanol production and lowers cost, if ethanol tolerance is not limiting as previously discussed. Increasing plant size will also significantly decrease ethanol cost (9.6%), and since this is a conservative estimate, the cost reduction may even be more, as discussed in Section 7.4.1. Deviation of the water recycle rate from the base case value effects ethanol cost because of the cost for makeup water and the xylose lost in the unreturned stream.

Significant cost reductions can also be achieved by striving for the highest onstream time or reliability. Finally, an analysis of the regression equations for SSF ethanol yield versus cellulase loading indicates that a reduction in ethanol cost is achieved by operating at a higher cellulase loading, thus generating a higher ethanol yield. Again, this assumes that ethanol tolerance is not a problem.

An analysis was conducted in which process variables are simultaneously varied to determine the compound effect of many process improvements. The results, shown in Section 7.5, gave an ethanol cost of 66.5¢/gal. This shows the potential cost of ethanol if all the improvements can be made. However, this is not an inclusive list. There is room for improvements in other areas that may reduce the cost even more or offset some of the more optimistic assumptions made in the current list. Furthermore, changes in the technology resulting in new designs and processes may reduce the cost of ethanol from biomass even further.

## Section 9.0

### Recommendations for Future Work

#### 9.1 Work to Strengthen Understanding of the Base Case Process

##### 9.1.1 In-house Integration Research

There is a lack of adequate information on the performance of each processing step using hydrolyzates and material produced from the previous steps. Previous work has not used hydrolyzates that have been systematically carried through the steps of xylose fermentation, cellulase production, and SSF. The experimental data used in this study were not obtained under the actual conditions of the process. Thus, the primary goal of integration research is to verify performance parameters such as yield and reaction rates, using the substrates and operating conditions described in the base case analysis. The following is a list of work and questions identified in this study that need to be answered by the integration effort.

**9.1.1.1 Yields and Rates.** Ethanol yields and rates for SSF, xylose fermentation, and cellulase production must be obtained for substrates described in the base case. Thus, pretreated high-solids hydrolyzates must be generated and carried sequentially through the steps of neutralization, xylose fermentation, cellulase production, and SSF. This work must consider the effect of fermenting a stream that contains gypsum, xylose, lignocellulosic particles, other potentially toxic by-products from prehydrolysis, and, in the case of SSF, microorganisms and fermentation products from the previous fermentation.

**9.1.1.2 Seed Cultures.** The growth of seed cultures on hydrolyzates to adequate cell concentrations must be verified and measured and shown not to affect ethanol production in both xylose fermentation and SSF. In particular, the growth of yeast culture for SSF on cellulose being simultaneously hydrolyzed to glucose must be tested, and cell concentrations and growth rates must be measured and optimized for SSF. Additionally, seed culture growth of *T. reesei* on hydrolyzate also needs to be tested and shown to not affect cellulase production.

**9.1.1.3 Nutrients.** SSF and xylose fermentation are assumed to require no supplemental nutrients above what is contained in the recycled water. However, as shown in Appendix A-6, if nutrients are required they will contribute significantly to the cost of ethanol. Although nutrients might not be required at the levels used in experimental work from which yield and performance data is taken, some supplemental nutrients might still be needed even with the use of recycled water. This supplemental nutrient requirement must be determined for both seed culture growth and the anaerobic fermentations, as they may be different.

**9.1.1.4 Base and Antifoam Consumption.** Although base usage is not accurately known for cellulase production, it is a small fraction of the cost (2.4%). The majority of the base cost is for xylose fermentation. The base usage is determined from experimental data, but for pure xylose solutions with a complex media. The actual base usage is needed for xylose fermentation of prehydrolyzed material with a minimal media.

Although antifoam cost is not significant, it is not known that corn oil will be a suitable foam suppressant. This fact should be verified or a suitable low-cost alternative should be found.

**9.1.1.5 Material Balance Closure.** Cell yield coefficients (g cells/g substrate) and optimum inoculum levels are unknown for this process. This study assumed cell yields of 0.5 g cells/g substrate; however, it is important to determine the actual values because this represents substrate that could be converted to ethanol. Likewise, the inoculum procedure and volumes followed current experimental practice that may

not be the optimum for the process. Seed growth rates and cell concentrations need to be determined for optimum ethanol production.

Along with ethanol yields, it is necessary to determine other by-product yields. Glycerol, acetaldehyde, and fusel oils are produced during yeast fermentations, but these have never been quantified for SSF. Products such as xylitol, glycerol, and xylulose have been measured for the *E. coli* xylose fermentations, but not on hydrolyzates with minimal media. More accurate and detailed product information is necessary to complete overall material balances for the process.

Accurate information is needed on vent gas composition for both xylose fermentation and SSF. This is necessary so that an optimized ethanol recovery system can be designed.

**9.1.1.6 Process Water Recycle Stream.** The process uses a liquid recycle stream from the lignin centrifuge to both downsize the waste treatment system and return nutrients back to the fermentations. However, this also returns to the process by-products that may build up to inhibitory levels. Once ethanol yields have been determined without recycle, the effect of liquid recycle on the fermentations can be investigated.

### **9.1.2 Subcontracted Equipment Testing Program**

Some of the assumptions made in this study can only be answered by vendor equipment testing. Critical equipment issues and concerns follow.

**9.1.2.1 Milling.** Little test data are available on the power requirements or chip characteristics produced by a disk refiner or other milling devices. Disk refining produces a chip that is more elongated than a chip produced by a hammer mill, which may then hydrolyze at a different rate. The power requirements used in this study was an estimate from the manufacturer (ABB Sprout-Bauer 1990). Testing is required to determine the exact power requirements for a variety of mill types and the performance of milled material during prehydrolysis and SSF, and to determine their handling characteristics.

**9.1.2.2 Prehydrolysis.** The impregnation and prehydrolysis reactors in the base case are continuous pulp digesters manufactured by Black and Clawson, Inc. Although pulp digestion is similar to the prehydrolysis step, the performance of this equipment with the conditions used in this study still needs to be verified. In particular, it must be shown that chips can be fully impregnated with acid in 10 min and that xylose yields obtained on the bench scale can be achieved by the large-scale equipment.

**9.1.2.3 Solids Concentration.** This study assumed that the solids concentration of the prehydrolyzed slurry pumped from the bottom of the blowdown tank is 12%. This assumption is critical for sizing of all downstream fermenters and could significantly increase or decrease the capital cost and power requirement for xylose fermentation, SSF, and ethanol purification. The actual solids concentration that could be pumped needs to be determined.

**9.1.2.4 Mixing Power for Xylose Conversion and SSF.** Mixing power requirements for both aerobic and anaerobic fermentations are required in xylose fermentation and SSF. This is particularly important in SSF where approximately 10% of the total plant power use is required for mixing in the fermenters. Some mixing is needed to promote transport of glucose away from the cellulose particles (Elander 1988), thus maintaining hydrolysis rates. However, the power requirements are unknown (0.1 hp/1,000 gal is assumed in this study). Because the current design is based on continuous flow through the fermenters, no mixing may be required, or more mixing may be required because of the necessity for transport and suspension of the solids. Large-scale vendor testing and advice is required to gain a better understanding of this issue.

**9.1.2.5 Mixing Power and Aeration for Aerobic Fermentations.** Along with mixing power, aeration requirements are also needed for cellulase production and seed growth in xylose fermentation and SSF. Each area is a major consumer of compressed air, but SSF is the largest user because of the large seed fermenter volumes. The air compressors are the major power user in the utilities section of the plant; therefore, it is important to quantify the fermenter air requirements. Oxygen uptake requirements as a function of time are needed for all four seed cultures (*T. reesei*, *E. coli*, *S. cerevisiae*, and *B. clausenii*) and in the cellulase production fermenter. In this study, assumptions were made for aeration requirements. Large-scale vendor testing and advice is needed to gain a better understanding of these issues.

**9.1.2.6 Fermentation Heat.** Heat of fermentation data is required for all three main fermentations including seed culture growth. In this study, metabolic heat production by *T. reesei* is based on a correlation with oxygen uptake, but oxygen uptake is not accurately known and the correlation may not be applicable to fungi systems. For xylose fermentation and SSF, metabolic heat production is assumed to be small compared to energy loss during the conversion of cellulose/xylose to ethanol. Heat production is important because the primary cooling loads for the entire plant are from removal of heat from the fermenters. Vendor testing in Sections 9.1.2.4 and 9.1.2.6 will provide data to make better heat balance calculations.

**9.1.2.7 Distillation Column.** The technology for ethanol distillation at high concentrations (4%-10%) is well developed and thus does not require any research work. However, in this study, the column is assumed to be capable of handling a particulate slurry. This operation is common for the corn-to-ethanol industry where the dilute ethanol stream contains residual fibers and protein from the corn kernels. Testing is required of an appropriate column to verify operation with lignocellulosic particles at the expected solid concentrations.

**9.1.2.8 Lignin Separation.** The bottom product from the distillation column contains lignin particles, unreacted cellulose and xylan, and cells that must be separated from the liquid and sent to the boiler. Vendor testing is required to determine the best method of separating this material and obtaining data on solid concentration and percent recovery of the dewatered material.

**9.1.2.9 Boiler.** The operation of the boiler and drying system must be assessed on actual feed material, particularly with regard to operation in the presence of gypsum. It is assumed that gypsum would not pose a problem for the boiler, so a gypsum separation system is not included in this study.

**9.1.2.10 Heat Exchange Equipment Testing.** Because of the high solids content of many of the streams that pass through heat exchangers, it is important to determine the best type of exchanger for each service. Vendor testing can provide much of the information required.

**9.1.2.11 Waste Treatment.** In this study, the waste treatment system consisted of an anaerobic digestion process followed by aerobic polishing as designed by Badger Engineers, Inc. (1984) for a biomass-to-ethanol plant using dilute sulfuric acid hydrolysis. The enzymatic process will produce a different waste stream that may not require such an extensive waste treatment system. A representative waste stream is required that could be tested to determine the best waste treatment system.

### **9.1.3 Subcontracted Work to Commercial Engineering Company**

In this study, carbon steel has been assumed to be adequate for most of the equipment needs, except in areas that have an obvious need for special alloys (e.g., pretreatment). The biggest assumption in this regard is the use of carbon steel fermenters in both xylose fermentation and SSF. Although current industry experience is with stainless steel fermenters, there are few data on the performance of carbon steel or alternative materials with aerobic or anaerobic fermentations. The assumption of a low-cost material



for the SSF fermenters is important because of the large fermentation capacity necessitated by the long fermentation time (7 days). It is necessary to get advice on materials of construction issues from a commercial engineering company.

#### **9.1.4 In-house Pilot Plant Operations**

Many issues surrounding the continuous long-term operation of an integrated large-scale base case process can only be answered by conducting pilot scale operations. Foremost among these will be demonstrating and verifying repeatable process performance on a large scale. Also, information on plant and equipment reliability, availability, and maintainability (RAM) on a long-term continuous service basis must be obtained from pilot operations. Enough data must be obtained to allow accurate and reliable design of a demonstration commercial plant.

Some of the specific issues that will be addressed with a pilot operation are given below:

- Large-scale pretreatment performance over an extended operating time
- Large-scale fermenter performance over an extended operating time. This includes yields, rates, mixing power, aeration rates, fermenter heat duties, etc.
- Equipment performance over an extended period of time
- Possible gypsum plating on critical process equipment. Gypsum is produced during neutralization of the sulfuric acid. Because the amount produced is greater than the saturation concentration, it is carried along in the process in both the soluble and insoluble forms. This is not expected to be a problem for most of the process because the gypsum is produced at 100°C and has increasing solubility with decreasing temperature. However, in the beer column reboiler, the solution is again heated to slightly greater than 100°C, which may cause precipitation and plating out of the gypsum. The plating of gypsum needs to be considered for the entire process, but especially in the beer column reboiler.
- Effect of recycled process water on long-term performance of system
- Material of construction issues as already noted in Section 9.1.3
- Large-scale process controllability verified
- A CIP/CS system appropriate for this process determined. Although a CIP/CS system is included in the current study, the exact requirements for such a system are unknown. Operating experience with pilot scale equipment will determine the requirements for this system.
- Large-scale safety of operations verified.

#### **9.2 Work to Develop New Technologies for Improving the Base Case Process**

Efforts in basic research and development should include working toward yield improvements, higher productivities, improved organisms, and new process designs. Material balance closure needs to be a priority for the experiments carried out in basic research. A list of key issues for basic research is given below in approximate order of priority:

1. Increase SSF yields to 90% without significantly decreasing other base case yields or increasing base case plant capital and operating costs.
2. Decrease SSF fermentation time to 3 days while keeping SSF yields at 90% and without significantly decreasing other base case yields or increasing base case plant capital and operating costs.
3. Decrease the amount of substrate for SSF and xylose seed fermenters without decreasing base case yields for SSF, xylose fermentations, or cellulase production, and without increasing base case capital and operating costs.
4. Decrease the amount of substrate to cellulase production without decreasing base case yields for SSF or xylose fermentation and without increasing base case capital and operating costs. Making cellulase from biomass feedstocks results in ethanol yield losses and capital expenditures that rise with increasing enzyme loadings to SSF. It is important to remember that support utilities such as fermentation air and chilled water must be included in the total cost. Enzymes with higher specific activity or a means to recycle enzymes could bring down the cost of cellulase production substantially.
5. Decrease use of power for milling without decreasing base case plant yields or increasing plant capital and operating costs.
6. Decrease the use of aeration to cellulase production and xylose and SSF seed fermenters without decreasing the base case plant yield or increasing base case plant capital or operating costs.
7. Minimize use of chilled water to cellulase production unit and vent gas condensers without decreasing base case plant yields or increasing base case capital and operating costs.
8. Improve ability to run all operations at a high solids content without jeopardizing base case yield and capital and operating costs.
9. Improve the ethanol tolerance of the microorganisms so that high solids operations can be achieved. Several of the sensitivity analyses in Section 7.0 of this report pointed out the importance of improving the ethanol tolerance of the SSF microorganisms and enzymes. Increasing solids concentration, improving ethanol yields, and using higher carbohydrate feedstocks can all lead to substantial reductions in ethanol cost. However, the positive effects of these improvements are diminished if ethanol tolerance limits the maximum ethanol concentration in SSF.

Other issues that may assume importance if certain assumptions in the base case are shown to be too liberal are:

- Nutrient requirements for xylose fermentation and SSF. Minimum nutrient requirements must be determined for the microorganisms under consideration. Nutrient cost could be a major operating cost for the process, and ultimately determine if a particular microorganism could be used in the process.
- Conversion of xylan to xylose during pretreatment. High overall xylan conversion to ethanol is required to reduce ethanol cost to competitive levels. If it is possible to recycle a large fraction of the process water, then xylan conversion during pretreatment is not quite as important. This is true only if xylanases can convert a significant fraction of the xylan to xylose during SSF. There will

always be a process water purge to balance the moisture entering with the feed, so some xylan and xylose would be lost even at maximum recycle rates.

Typically, there are many possible technical solutions to these issues. Members of the ethanol program research community should submit their ideas and from these the ones with the most potential should be selected.

### **9.3 Process Analysis**

Process analysis provides a tool by which changes and improvements in the technology can be assessed in terms of the process economics. For example, improvements in yields, productivities, or power requirements that result in significant cost reductions are particular areas in which to focus future research efforts. As such, process analysis is a valuable tool for determining the best directions and areas for future work. The design presented in this study is our best design based on current technology. As future research improves the technology, it will be necessary to continually analyze and update the process design and economics. This will provide the structured framework necessary to guide the process development from bench scale research to commercial operation.

Besides identifying areas for research, process analysis also provides goals for plant design in such areas as plant size, feedstock composition, and cost. As plant size is increased, economies of scale decrease ethanol cost. However, this benefit decreases because at some point it is no longer possible to scale up equipment. Instead, a doubling of plant capacity requires two separate flow trains that eliminate the benefits of economies of scale. In addition, average transportation costs for feedstock will probably rise with plant size due to a larger collection area. Process analysis will help to determine optimum plant capacities.

## Section 10.0

### References

- ABB Sprout-Bauer, Inc. 1990, R. Rodarmel, Personal Communication.
- A.D. Little, 1984, "Technical and Economic Feasibility of Enzyme Hydrolysis for Ethanol Production From Wood," Prepared for New York State Energy Research and Development Authority and Solar Energy Research Institute. Cambridge, MA: A.D. Little.
- Atkinson, B., and F. Mavituna, 1983, *Biochemical Engineering and Biotechnology Handbook*, New York: Stockton Press.
- Badger Engineers Inc., 1984, *Economic Feasibility Study of an Acid Hydrolysis Based Ethanol Plant*, Prepared for the Solar Energy Research Institute, Subcontract ZX-3-030-96-2, Cambridge, MA: Badger Engineers, Inc.
- Bevernitz, K.J., S.J. Gracheck, D.B. Rivers, D.K. Becker, K.F. Kaupisch, and G.H. Emert, 1982, "Development of Enzyme-Catalyzed Cellulose Hydrolysis Process For Ethanol Production," *Energy from Biomass and Waste VI*, ed. Donald L. Klass, Chicago, IL: Institute of Gas Technology, pp. 897-918.
- Brownell, H.H., and J.N. Saddler, 1984, "Steam-Explosion Pretreatment for Enzymatic Hydrolysis," *Biotech. Bioeng. Symp.* No. 14, pp. 55-68.
- Callens, M., H. Kersters-Hilderson, O. Van Opstal, and C.K. De Bruyne, 1985, "D-Xylose Isomerase from *Streptomyces violaceus-niger*: Optimization of Fermentation Conditions," *Biot. Letters*, Vol. 7, No. 8, pp. 597-600.
- Chem Systems, Inc., 1984, *Economic Feasibility Study of an Enzymatic Hydrolysis Based Ethanol Plant With Prehydrolysis Pretreatment*, Prepared for the Solar Energy Research Institute, SERI Subcontract XX-3-03097-2, Tarrytown, NY: Chem Systems, Inc.
- Chem Systems, Inc., 1989, *Assessment of Cost of Methanol From Biomass*, SERI Subcontract Report.
- Chem Systems, Inc., 1990, *Technical and Economic Evaluation: Wood to Ethanol Process*, SERI Subcontract Report.
- Chemical Marketing Reporter*, Nov. 27, 1989.
- Chemical Marketing Reporter*, March 26, 1990.
- Chiang, L., H. Hsiao, P. Ueng, L. Chen, and G. Tsao, 1981, "Ethanol Production from Xylose by Enzymatic Isomerization and Yeast Fermentation," *Biotech. Bioeng. Symp.* No. 11, pp. 263-274.
- COADE, 1983, *CHEMCOST Capital Cost and Profitability Analysis (Program for Microcomputers)*, Houston, TX: COADE, Inc.

- Deshparde, V., H.S. Roman, and M. Rao, 1983, "Simultaneous Saccharification and Fermentation of Cellulose to Ethanol Using *Penicillium funiculosum* Cellulase and Free or Immobilized *Saccharomyces uvarum* Cells," *Biotech. Bioeng.*, Vol. 25, pp. 1679-1684.
- Durand, H., M. Clanet, and G. Tiraby, 1988, "Genetic Improvement of *Trichoderma reesei* for Large Scale Cellulase Production," *Enz. Microb. Technol.*, Vol. 10, pp. 341-346.
- Elander, R.T., 1988, "Mixing Requirements for Enzymatic Hydrolysis for Cellulose," Master's Thesis, Colorado State University, Fort Collins, CO.
- Genencor Cellulase 150L Product Sheet, Genencor, Inc., South San Francisco, CA.
- Ghosh, P., N.B. Pamment, and W.R.B. Martin, 1982, "Simultaneous Saccharification and Fermentation of Cellulose: Effect of  $\beta$ -D-Glucosidase Activity and Ethanol Inhibition of Cellulases," *Enzyme Microb. Technol.*, Vol. 4, pp. 425-430.
- Gong, C., L. Chen, M. Flickinger, L. Chiang, and G. Tsao, 1981, "Production of Ethanol from D-Xylose by Using D-Xylose Isomerase and Yeast," *Appl. Environ. Microb.*, Vol. 41, No. 2, pp. 430-436.
- Grohmann, K., R. Torget, and M. Himmel, 1985, "Optimization of Dilute Acid Pretreatment of Biomass," *Biotech. Bioeng. Symp.* No. 15, pp. 59-80.
- Grohmann, K., R. Torget, and M. Himmel, 1986, "Dilute Acid Pretreatment of Biomass at High Solids Concentrations," *Biotech. Bioeng. Symp.* No. 17, pp. 137-151.
- Guthrie, K.M., 1974, *Process Plant Estimating and Control*, Solana Beach, CA: Craftsman Book Company of America.
- Hahn-Hagerdal, B., S. Berner, and K. Skoag, 1986, "Improved Ethanol Production from Xylose with Glucose Isomerase and *Saccharomyces cerevisiae* Using the Respiratory Inhibitor Azide," *Appl. Microb. Biotech.*, Vol. 24, pp. 287-293.
- Hall, R.H., J. Matley, and K.J. McNaughton, April 5, 1982, "Current Cost of Process Equipment," *Chemical Engineering*, Vol. 89, No. 7, pp. 80-116.
- Hall, R.H., W.M. Vataavuk, and J. Matley, Nov. 21, 1988, "Estimating Process Equipment Cost," *Chemical Engineering*, Vol. 95, No. 17, pp. 66-75.
- Hendy, N., C. Wilke, and H. Blanch, 1982, "Enhanced Cellulase Production Using Solka Floc in a Fed-Batch Fermentation," *Biot. Letters*, Vol. 4, No. 12, pp. 785-788.
- Hendy, N.A., C.R. Wilke, and H.W. Blanch, 1984, "Enhanced Cellulase Production in Fed-Batch Culture of *Trichoderma reesei* C30," *Enz. Microb. Technol.*, Vol. 6, pp. 341-346.
- Himmel, M., M. Tucker, J. Baker, C. Rivard, K. Oh, and K. Grohmann, 1986, "Comminution of Biomass: Hammer and Knife Mills," *Biot. and Bioeng. Symp.* No. 15, pp. 39-58.
- Himmelblau, D., 1962, *Basic Principles and Calculations in Chemical Engineering*, Englewood Cliffs, NJ: Prentice-Hall.

- Hinman, N.D., J.D. Wright, W. Hoagland, and C.E. Wyman, 1989, *Appl. Biochem. Biotech.*, Vol. 20/21, p. 391.
- Huang, S.Y., and J.C. Chen, 1988, "Ethanol Production in Simultaneous Saccharification and Fermentation of Cellulose With Temperature Profiling," *J. Ferment. Technol.*, Vol. 66, No. 5, pp. 509-516.
- Icarus Corp. July 1985. *Icarus Cost System Report*.
- Icarus Corp. April 1987. *Icarus Cost System Report*.
- Ingram, L.O., T. Conway, D.P. Clark, G.W. Sewell, and J.F. Preston, 1987, "Genetic Engineering of Ethanol Production in *Escherichia coli*," *Appl. Environ. Microb.*, Vol. 53, No. 10, pp. 2420-2425.
- Ingram, L.O., and T. Conway, 1988, "Expression of Different Levels of Ethanologenic Enzymes from *Zymomonas mobilis* in Recombinant Strains of *Escherichia coli*," *Appl. Environ. Microb.*, Vol. 54, No. 2, pp. 397-404.
- Kim, S.B., and Y.Y. Lee, 1987, "Kinetics in Acid-Catalyzed Hydrolysis of Hardwood Hemicellulose," *Biotech. Bioeng. Symp. No. 17*, p. 71.
- Krull, D., Nov. 27, 1989, Personal Communication, Novo Laboratories, Inc.
- Kwarteng, I.K., 1983, "Kinetics of Acid Hydrolysis of Hardwood in a Continuous Plug Flow Reactor," Ph.D Thesis, Dartmouth College, Hanover, NH.
- Lastick, S.M., 1989, Personal Communication, Solar Energy Research Institute.
- Lastick, S.M., M.Y. Tucker, V. Machedonski, and K. Grohmann, 1986, "Overproduction of *E. coli* Xylose Isomerase," *Biot. Letters*, Vol. 8, No. 1, pp. 1-6.
- Lastick, S.M., M.Y. Tucker, J.R. Beyette, G.R. Noll, and K. Grohmann, 1989, "Simultaneous Fermentation and Isomerization of Xylose," *Appl. Microb. Biotech.*, Vol. 30, pp. 574-579.
- Lastick, S.M., A. Mohagheghi, M.P. Tucker, and K. Grohmann, 1990, "Simultaneous Fermentation and Isomerization of Xylose to Ethanol at High Xylose Concentrations," *Appl. Biochem. Biotech.*, Vol. 24/25, pp. 431-440.
- Mackie, K.L. et al., 1985, *J. Wood Chemistry*, Vol. 5, pp. 405-425.
- McKibbins, S.W., J.F. Harris, J.F. Saeman, and W.K.I. Neill, 1962, "Kinetics of Acid Catalyzed Conversion of Glucose to 5-Hydroxymethyl-2-Furaldehyde and Levulinic Acid," *Forest Products Journal*, Vol. 12, pp. 17-23.
- McLean, D., and M.F. Podrutzny, 1985, "Further Support for Fed-Batch Production of Cellulases," *Biot. Letters*, Vol. 7, No. 9, pp. 683-688.
- Means, G.E., and O.D. Monera, 1989, *Immobilization of E. coli Xylose Isomerase*, Draft report prepared for the Solar Energy Research Institute, Subcontract No. XL-8-07164-1.
- Meyers, S.G., 1978, "Ethanollic Fermentation During Enzymatic Hydrolysis of Cellulose," *AIChE Symp. Ser.*, Vol. 74, No. 181, pp. 79-84.

- Mohagheghi, A., K. Grohmann, and C.E. Wyman, 1988, "Production of Cellulase on Mixtures of Xylose and Cellulose," *Appl. Biochem. Biotech.*, Vol. 17, pp. 263-278.
- Morbark Company Catalog, 1983.
- Orton, D., A. Protopapas, and M. Tsiakkas, 1988, "Experimentation and Process Design for the Fermentation and Isomerization of Xylose," SERI Report RR-231-3211, Golden, CO: Solar Energy Research Institute.
- Perry, R.H., and C.H. Chilton, 1973, *Chemical Engineers' Handbook*, 5th Ed., New York: McGraw-Hill.
- Peters, M.S., and K.D. Timmerhaus, 1980, *Plant Design and Economics for Chemical Engineers*, 3rd edition, New York: McGraw-Hill.
- Rivard, C., 1990, Personal Communication, Solar Energy Research Institute.
- Ryu, D., R. Andreotti, M. Mandels, B. Gallo, and E.T. Reese, 1979, "Studies on Quantitative Physiology of *Trichoderma reesei* with Two-Stage Continuous Culture for Cellulase Production," *Biotech. Bioeng.*, Vol. 21, p. 1887.
- Seaman, J.F., 1945, "Kinetics of Wood Saccharification," *Industrial and Engineering Chemistry*, Vol. 37, p. 43.
- Schell, D.J., N.D. Hinman, C.E. Wyman, and P.J. Werdene, 1990, "Simultaneous Saccharification and Fermentation Using Whole Cellulase Culture Broth," *Appl. Biochem. Biotech.*, Vol.24/25, pp. 287-298.
- Schell, D.J., 1990, Solar Energy Research Institute, Personal Communication.
- Schellenburg, G.D., A. Sarthy, A.E. Larson, M.P. Backer, J.W. Crabb, M. Lidstrom, B.D. Hall, and C.E. Furlong, 1983, "Xylose Isomerase from *Escherichia coli*," *J. Biol. Chem.*, Vol. 259, No. 11, pp. 6826-6832.
- Schwald, W., C. Breuil, H. Brownell, M. Chan, and J. Saddler, 1989, "Assessment of Pretreatment Conditions to Obtain Fast Complete Hydrolysis on High Substrate Concentrations," *Appl. Biochem. Biotech.*, Vol. 20/21, pp. 29-44.
- Shafizadeh, F., 1984, "The Chemistry of Pyrolysis and Combustion," *The Chemistry of Solid Wood*, edited by R.M. Rowell, Washington, D.C.: American Chemical Society, pp. 489-530.
- Sheir-Ness, G., and B.S. Montecourt, 1984, "Characterization of the Secreted Cellulases of *Trichoderma reesei* Wild Type and Mutants During Controlled Fermentations," *Appl. Microbiol. Biotech.*, Vol. 26, pp. 46-53.
- Shoemaker, S.P., J.C. Raymond, and R. Bruner, 1981, "Cellulases: Diversity Amongst *Trichoderma* Strains," *Trends in the Biology of Fermentations for Fuels and Chemicals*, ed. by A. Hollaender et al., New York: Plenum Press.
- Skoog, K., and B. Hahn-Hagerdal, 1988, "Xylose Fermentation," *Enzyme Microb. Technol.*, Vol. 10, pp. 66-80.
- Spencer, T., 1989, Personal Communication, Virginia Polytechnic Institute.

Spindler, D.D., C.E. Wyman, A. Mohagheghi, and K. Grohmann, 1988, "Thermotolerant Yeast for Simultaneous Saccharification and Fermentation of Cellulose to Ethanol," *Appl. Biochem. Biotech.*, Vol. 17, pp. 279-293.

Spindler, D.D., C.E. Wyman, and K. Grohmann, 1989a, Evaluation of Thermotolerant Yeasts in Controlled Simultaneous Saccharification and Fermentation of Cellulose to Ethanol," *Biotech. Bioeng.*, Vol. 34, pp. 189-195.

Spindler, D.D., C.E. Wyman, K. Grohmann, and A. Mohagheghi, 1989b, "Simultaneous Saccharification and Fermentation of Pretreated Wheat Straw to Ethanol with Selected Yeast Strains and  $\beta$ -Glucosidase Supplementation," *Appl. Biochem. Biotech.*, Vol. 20/21, pp. 529-540.

Spindler, D.D., 1989, Personal Communication, Solar Energy Research Institute.

Sternberg, D., and S. Dorval, 1979, "Cellulase Production and Ammonia Metabolism in *Trichoderma reesei* on High Levels of Cellulose," *Biot. Bioeng.*, Vol. 21, pp. 181-191.

Stavis, P.E., and N.W.Y. Ho, 1985, "Overproduction of D-Xylose Isomerase in *Escherichia coli* by Cloning the D-Xylose Isomerase Gene," *Enzyme Microb. Technol.*, Vol. 7, pp. 592-596.

Stone and Webster Engineering Corp., 1985, *Economic Feasibility Study of an Enzyme-Based Ethanol Plant*, SERI Subcontract No. ZX-3-03097-1, Boston, MA: Stone and Webster Engineering Corp.

Strehler, A. -1987, "Handling and Storage of Straw and Woodchips," *Biomass Energy: From Harvesting to Storage*, New York: Elsevier, pp. 191-199.

Szczodrak, J., 1988, "The Enzymatic Hydrolysis and Fermentation of Pretreated Wheat Straw to Ethanol," *Biotech. Bioeng.*, Vol. 32, pp. 771-776.

Szczodrak, J., 1989, "The Use of Cellulases From a  $\beta$ -Glucosidase-Hyperproducing Mutant of *Trichoderma reesei* in Simultaneous Saccharification and Fermentation of Wheat Straw," *Biotech. Bioeng.*, Vol. 33, pp. 1112-1116.

Takagi, M., S. Abe, S. Suzuki, G.H. Emert, and N. Yata, 1977, "A Method for Production of Alcohol Directly from Cellulose Using Cellulase and Yeast," *Proc. Bioconversion Symp.*, Delhi: IIT, pp. 551-571.

Tangnu, S.K., H.W. Blanch, and C.R. Wilke, 1981, "Enhanced Production of Cellulase, Hemicellulase, and  $\beta$ -Glucosidase by *Trichoderma reesei* (Rut C-30)," *Biotech. Bioeng.*, Vol. 23, pp. 1837-1849.

Torget, R., M. Himmel, J.D. Wright, and K. Grohmann, 1988, "Initial Design of a Dilute Sulfuric Acid Pretreatment Process for Aspen Wood Chips," *Appl. Biochem. and Biotech.*, Vol. 17, pp. 89-104.

Touloukian, Y.S., and E.H. Buyco, 1970, *Thermophysical Properties of Matter*, Vol. 5, New York: Plenum Press.

Ulrich, G.D., 1984, *A Guide to Chemical Engineering Process Design and Economics*, New York: John Wiley and Sons.

University of Arkansas, 1982, *Final Technical Report: Develop and Demonstrate the Cellulose Process*, SERI Subcontract XR-0-9248-1, Fayetteville, AK: University of Arkansas.



- Walas, S.M., 1988, *Chemical Process Equipment*, Boston: Butterworths Publishers.
- Wang, D.I.C., C.L. Cooney, A.L. Demain, P. Dunnill, A.E. Humphrey, and M.D. Lilly, 1979, *Fermentation and Enzyme Technology*, New York: John Wiley and Sons.
- Warzywoda, M., F. Chevron, V. Ferre, and J. Pourquie, 1983, "Pilot Scale Production of Cellulolytic Enzymes by *Trichoderma reesei*," *Biotech. Bioeng. Symp.* No. 13, pp. 577-580.
- Watson, T.G., and I. Nelligan, 1983, "Pilot Scale Production of Cellulase by *Trichoderma reesei* (Rut C-30)," *Biot. Letters*, Vol. 5, No. 1, pp. 25-28.
- Watson, T.G., I. Nelligan, and L. Lessing, 1984, "Cellulase Production by *Trichoderma reesei* (Rut-C30) in Fed-Batch Culture," *Biot. Letters*, Vol. 6, No. 10, pp. 667-672.
- Wayman, M., and S.R. Parekh, 1988, "SO<sub>2</sub> Prehydrolysis for High Yield Ethanol Production from Biomass," *Appl. Biochem. Biotech.*, Vol. 17, pp. 33-44.
- Weast, R.C., ed., 1972, *Handbook of Physics and Chemistry*, Cleveland, OH: CRC Press.
- Wenzel, H., 1970, *The Chemical Technology of Wood*, New York: Academic Press.
- Wilke, C.R., and H.W. Blanch, 1985, *Process Development Studies on the Bioconversion of Cellulose and Production of Ethanol*, Annual Report to Solar Energy Research Institute, Golden, CO.
- Wocvha, M.G., D.L. Steuerwald, and K.E. Brooks, 1983, "Amplification of D-Xylose and D-Glucose Isomerase Activities in *Escherichia coli* by Gene Cloning," *Appl. Environ. Microb.*, Vol. 45, No. 4, pp. 1402-1404.
- Wright, J.D., and C.G. d'Agincourt, 1984, *Evaluation of Sulfuric Acid Hydrolysis Processes for Alcohol Fuel Production*, SERI/TR-231-2074, Golden, CO: Solar Energy Research Institute.
- Wright, J.D., C.E. Wyman, and K. Grohmann, 1988, "Simultaneous Saccharification and Fermentation of Cellulose: Process Evaluation," *Appl. Biochem. Biotech.*, Vol. 18, pp. 75-90.
- Wright, J.D., 1988, "Ethanol from Biomass by Enzymatic Hydrolysis," *Chemical Engineering Progress*, Vol. 84, No. 8, pp. 62-74.
- Yaws, C.L., 1977, *Physical Properties*, New York: McGraw-Hill.

## Appendix A

### Tradeoff Studies and Options Considered

#### Appendix A-1—Milling

The three most common techniques for biomass size reduction are hammer milling, knife milling, and disk refining. The power consumption for both hammer and knife milling of wheat straw, aspen, corn cobs, and corn stover has been studied by Himmel et al. (1986). This study determined that knife milling has the lowest power consumption and is the preferred technique. Thus, knife-milled material has been used for most of SERI's experimental work (Spindler et al. 1989a, Spindler et al. 1988).

However, based on the recommendations of ABB Sprout-Bauer, Inc. (1990), a manufacturer of all three types of mills, a disk refiner was chosen for this study. Both the hammer mills and the knife mills currently manufactured have limited capacity. For example, the capacity of the largest knife mill available is 8000 lb/h; thus, 20 mills are required to handle the same capacity as four disk refiners. The cost of 20 knife mills is \$3.1 MM compared to the cost of \$1.48 MM for four disk refiners. However, the greater power requirements for disk refiners (128 hp-h/ton, [ABB Sprout-Bauer 1990] compared to 85 hp-h/ton for knife milling [Himmel et al. 1986]) makes the total cost (operating plus capital charges) for each option approximately equal. But disk refiners were chosen because less solids handling equipment is required to feed four disk refiners when compared to the equipment required to feed 20 knife mills. Maintenance requirements are also expected to be less severe for disk refiners (knife mills would require frequent replacement of the knife blades).

#### Appendix A-2—Pretreatment

##### Introduction

Of the many biomass feedstocks available for ethanol production, one of the most abundant and cheapest is cellulosic biomass. Cellulose, a polymer of glucose, can be broken down into glucose by enzymes and then converted to ethanol by yeast. However, hydrolysis of cellulose in raw cellulosic biomass is difficult. This has been attributed to the crystallinity of cellulose and the lignin-hemicellulose sheath that surrounds the cellulose. Thus, some form of pretreatment is necessary to disrupt the lignin-hemicellulose sheath and increase the susceptibility of the cellulose to enzymatic attack. Pretreatment can also hydrolyze hemicellulose to its individual sugar components. In the case of hardwoods and wheat straw, the hemicellulose is composed primarily of the five-carbon sugar xylose, a sugar that can also be converted to ethanol. The conversion of xylose to ethanol improves the overall economics of the cellulosic biomass-to-ethanol process (Hinman et al. 1989).

Several processes can be used for pretreating biomass including autohydrolysis steam explosion, steam explosion with an acid catalyst, dilute sulfuric acid hydrolysis, and the organosolv process. The dilute-acid process uses low concentrations of sulfuric acid at relatively low temperatures (160°C for 10 min) to achieve almost complete conversion of the hemicellulose xylans to xylose (Grohmann et al. 1986, Torget et al. 1988). However, prior to this step, particle size is reduced to nearly 1.0 mm, requiring significant amounts of energy. Both steam-explosion processes use high-pressure steam and rapid depressurization to reduce the size of the biomass particle and partially hydrolyze the hemicellulose fraction. Both require operationally complex steam-explosion guns. With autohydrolysis steam explosion, the yield of xylose is low (30%-50%) (Wright 1988). However, the yield can be improved by using a catalyst, such as  $\text{SO}_2$  (Schwald et al. 1989), which is the basis for the acid-catalyzed steam-explosion process. The organosolv process uses an organic solvent to dissolve the lignin and hemicellulose fractions from the cellulose. The lignin is then precipitated from solution, leaving the xylose in the liquid stream. This process is

complicated and expensive, but produces a high-quality lignin stream that could be converted to high-value products.

Because of the low xylose yields and their negative impact on ethanol production economics, a detailed analysis of autohydrolysis steam explosion was not carried out in this study. Furthermore, because of the expense and complexity of the organosolv process and because there are no current large markets for high-quality lignin, this process was also not considered further. The economics of the two remaining pretreatment options, steam explosion with an acid catalyst and dilute sulfuric acid pretreatment, were evaluated with two different feedstocks: wheat straw and aspen wood chips. These materials are representative of the performance expected for two of the most abundant categories of cellulosic biomass, herbaceous and wood energy crops.

## **Methodology**

Each of the four feedstock/pretreatment combinations is analyzed on the basis of total sugar (glucose as unconverted cellulose and xylose) produced, which could then be converted to ethanol. The economic information is summarized on the basis of total sugars, assuming 90% of the cellulose leaving the pretreatment process could be converted to glucose. However, the analysis does not consider conversion of cellulose to glucose or subsequent conversion of sugars to ethanol, but only considers the pretreatment processes. A different yield of glucose from cellulose will change the absolute cost of the processes but not the relative comparison between processes.

## **Process Description**

**Acid-Catalyzed Steam Explosion With Aspen Wood.** A flowsheet of the acid-catalyzed steam explosion process for aspen wood chips is shown in Figure A-1. The design for the steam-explosion system is adopted from a design by Stone and Webster Engineering Corp. (Stone and Webster Engineering Corp. 1985), and the design of the SO<sub>2</sub> recovery section is based on the work of several groups (Schwald et al. 1989, Mackie et al. 1985, Wayman and Parekh 1988, Brownell and Saddler 1984). Finally, the design of the lime slurry preparation system section is adopted from a study by Badger Engineers, Inc. (1984).

Aspen wood chips are stored in an open pile and delivered by front-end loaders to a screen that removes oversized material, which is sent to a rechipper. The screened material is loaded into a stainless-steel steam-explosion gun. Each stainless-steel gun is a 3.5 ft (1.07 m) diameter pipe, 14.5 ft (4.45 m) long, designed for 650 psig (4480 kPa), and sealed on each end by quick-opening, full-port plug or ball valves. Once the chips are sealed inside the gun, steam and vaporized sulfur dioxide are added, and the chips are cooked for 2 min. Sulfur dioxide added to the gun is taken as a liquid from a carbon steel storage vessel, pumped to 600 psig (4236 kPa), vaporized, and mixed with steam. After cooking, the material is blown down into a 316 stainless steel flash vessel operating at 60 psig (515 kPa). The steam-exploded wood then flows to a second 316 stainless-steel atmospheric flash tank where final cooling takes place.

Vapor from each of the flash tanks is sent to partial condensers (304 stainless-steel tubes/carbon steel shell). The condensate, still containing a small amount of SO<sub>2</sub>, is collected in a condensate receiver, then pumped to a 316 stainless-steel open-steam stripping column. The SO<sub>2</sub> removed from the top of the column is combined with the uncondensed SO<sub>2</sub> from the partial condensers, compressed to a liquid, and recycled back to the liquid SO<sub>2</sub> storage tank. The water from the bottom of the column is sent to waste treatment. This extensive SO<sub>2</sub> recovery system is employed to significantly reduce any discharge to the environment.



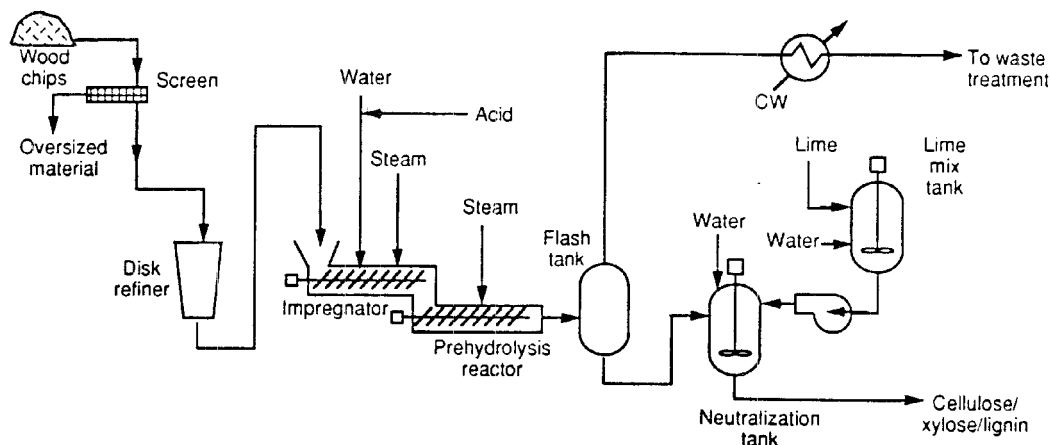


Figure A-2. Flowsheet for dilute sulfuric acid pretreatment of aspen wood chips

**Dilute Sulfuric Acid Pretreatment With Wheat Straw.** The design of the feed handling/storage and preparation areas is the same as that described for acid-catalyzed steam explosion with wheat straw. After screening the straw is sent to a disk refiner, and the remainder of the process is as described previously for dilute sulfuric acid pretreatment with aspen wood.

**Plant Capacity and Feed Composition.** Process flow diagrams for each feedstock were developed based on a delivered flow rate of 160,000 lb/h (72,700 kg/h) of dry feedstock to the pretreatment section of the plant. Aspen wood enters the plant at 50% moisture, and wheat straw enters the plant at 12% moisture (Strehler 1987). The chemical composition of each feedstock is given in Table A-1.

### Design Basis

**Acid-Catalyzed Steam Explosion.** The chips are held in the steam-explosion guns for 2.0 min at 240°C (Schwald et al. 1989). An additional 0.5 min is needed for loading and preheating, and another 0.5 min is needed for blowdown and cleaning. The total cycle time per gun is 3 min, although this has not been demonstrated at the large scale assumed for this plant.

The SO<sub>2</sub> used for steam explosion is 1.6 kg per 100 kg of dry wood, and the high-pressure steam use is 0.85 kg per kg dry wood (Schwald et al. 1989). For wheat straw, the SO<sub>2</sub> use is the same, and the high-pressure steam use is calculated to be approximately 0.45 kg per kg of dry straw because of the lower water content. The xylan conversion from a SO<sub>2</sub>-catalyzed steam explosion is 75% conversion to xylose, 15% conversion to furfural, 5% unchanged, and 5% degraded to solid products. Cellulose is assumed unchanged. The SO<sub>2</sub> is converted as follows: 9.2% to sulfuric acid, 9.7% to lignin sulfonic acids, 74.2% unconverted and available for recycle, and 6.5% retained with the lignin (Schwald et al. 1989). The washing filter belt system is a five-stage washing operation, similar to a paper pulp stock washer, in which 99% of the solubles are recovered (Stone and Webster Engineering Corp. 1985).

**Dilute Sulfuric Acid Pretreatment.** The milling step requires 125 hp/ton/h (94 kW/ton) of electrical power for wood and 12.5 hp/ton/h (6.4 kW/ton) for wheat straw (ABB Sprout-Bauer 1990). The impregnator operates at 100°C with a 10-min residence time (Torget et al. 1988). The prehydrolysis

**Table A-1. Chemical Composition of Aspen Wood and Wheat Straw**

	Aspen Wood (Badger Engineers 1984) (%)	Wheat Straw (Grohmann et al. 1986) (%)
Cellulose	46.2	40.8
Xylan	24.0	27.0
Lignin	24.0	18.4
Ash	0.2	11.2
Other	5.6	2.6

reactor operates at 160°C for a 10-min residence time and with an acid concentration of 1 wt % after steam and water addition (Torget et al. 1988). Xylan is assumed converted as follows: 80% to xylose, 13% to furfural, and 7% unconverted (Grohmann et al. 1986). Also, during prehydrolysis, 4% of the cellulose is converted to glucose (Seaman 1945); the rest is assumed unchanged.

### **Capital Cost Estimate and Economic Analysis**

Heat and material balances were developed and used to specify equipment sizes. Purchased equipment cost is estimated using information from COADE (1983), Icarus Corp. (1987), Guthrie (1974), Stone and Webster Engineering Corp. (1985), and Badger Engineers, Inc. (1984). Total fixed investment is estimated as 2.85 times the purchased equipment cost (Chem Systems, Inc. 1990) plus an additional 2.0% for miscellaneous equipment. Working capital is 4.8% and startup cost is 5.0% of total fixed investment (Chem Systems, Inc. 1990). The annual capital charge (depreciation, taxes, insurance, and rate of return) is total capital invested (fixed plus working plus startup cost) times a fixed charge rate (FCR) of 0.20, typical for these types of plants (Chem Systems, Inc. 1990, Chem Systems, Inc. 1989). Chemical costs are taken from the *Chemical Marketing Reporter* (1990). Utility costs for process water, cooling water, and steam are estimated from Peters and Timmerhaus (1980), and electricity is assumed to cost \$0.04/kWh. Manpower required is estimated from a previous study (Chem Systems, Inc. 1990) as 14 laborers at \$29,800/y and 3 foremen at \$34,000/y. Maintenance is 3.0% of total capital invested, and overhead is 65% of labor plus maintenance. Insurance and taxes are 1.5% of total fixed investment. By-product credit is taken for lignin sent to the boiler and is estimated as the heating value of lignin divided by the total heating value of the feedstock times the feedstock cost.

### **Results**

An economic summary of acid-catalyzed steam explosion for both aspen wood and wheat straw is given in Table A-2 for a feedstock cost of \$42/dry ton (Wright et al. 1988) and a FCR of 0.2. The sugar selling price (glucose and xylose) as a function of feedstock cost is shown in Figure A-3 for both feedstocks. It is somewhat cheaper to produce sugars from aspen wood for the same feedstock cost. This is due to the greater amount of sugars contained in aspen wood (70% cellulose and xylans) when compared to wheat straw (68% cellulose and xylans). Also, aspen wood has a larger by-product credit because of its larger lignin content.

An economic summary of dilute sulfuric acid pretreatment for both aspen wood and wheat straw is given in Table A-3 for a feedstock cost of \$42/dry ton. The sugar selling price as a function of feedstock cost is also shown in Figure A-3. In this case, aspen wood also results in a lower sugar selling price for the same feedstock cost, except below \$35/dry ton where the costs are approximately the same. In this case, the advantages of the higher carbohydrate content of the aspen wood are offset by the greater electrical cost required to mill the aspen.

**Table A-2. Economy Summary of the Steam Explosion Process**

Grassroots plant, first-quarter 1990 cost

Plant Capacity: 160,000 dry lb/h

Total Capital Investment:

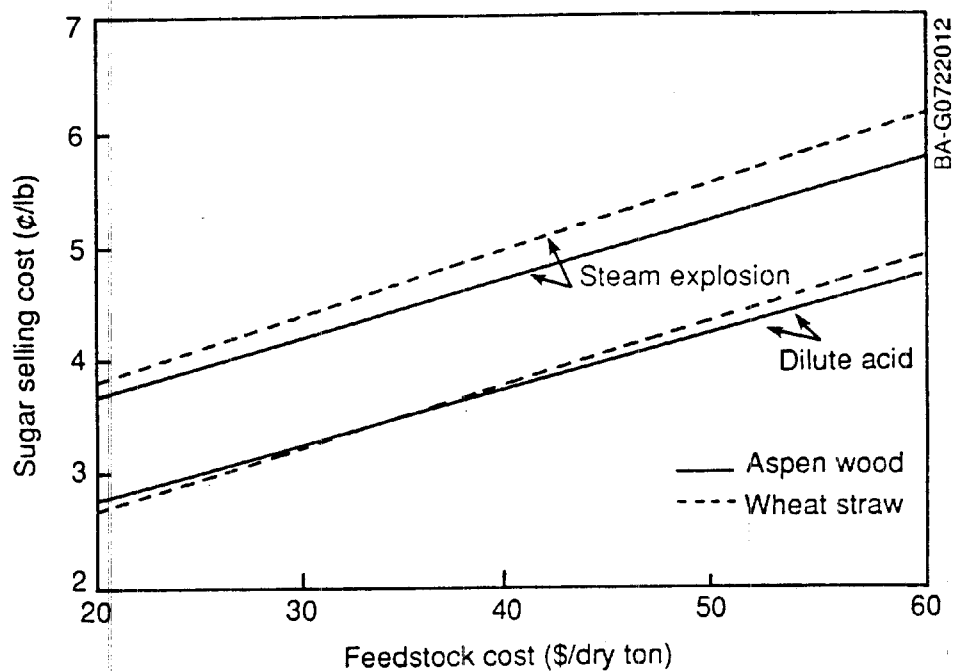
Aspen wood plant \$53.4 MM

Wheat straw plant \$54.4 MM

	Aspen	Wheat Straw
		¢/lb sugars
<b>Raw Materials</b>		
Feed	3.28	3.44
SO <sub>2</sub>	0.07	0.08
Lime	0.01	0.01
<b>Utilities</b>		
Process water	0.07	0.08
Cooling water	0.09	0.10
Steam-60 psig	0.28	0.26
Steam-600 psig	0.30	0.16
Electricity	0.03	0.05
Labor	0.06	0.06
Maintenance	0.18	0.19
Overhead	0.16	0.16
Insurance and Taxes	0.08	0.08
<b>By-product Credits</b>		
Lignin	1.06	0.96
<b>Capital Charges</b>	1.30	1.39
<b>Totals</b>	<b>4.85</b>	<b>5.11</b>

Feedstock Cost: \$42/dry ton

FCR: 0.20



**Figure A-3. Sugar selling cost as a function of feedstock type and cost for both aspen wood and wheat straw**

### Discussion

The choice of feedstocks for sugar production will probably be governed by plant location. Plants situated in heavy food crop agricultural areas will probably use agricultural residues such as wheat straw. In other areas, where agricultural residues are not readily available, wood energy crops will probably be used. There is not an overwhelming advantage to either feedstock, particularly for dilute sulfuric acid, if they are obtained at the same price. However, feedstock cost does have a significant effect on the selling price of sugar. A decrease in feedstock cost from \$60 to \$20 per dry ton decreases the selling price by approximately 2.0 ¢/lb sugar, which is a 35% reduction for steam explosion and a 42% reduction for dilute sulfuric acid.

The data in Tables A-2 and A-3 show that dilute sulfuric acid is approximately 20% cheaper than steam explosion using the process configuration and yields assumed in this study. The higher cost for steam explosion is primarily due to the higher capital cost associated with SO<sub>2</sub> recovery. These results are based on our current understanding of each of these processes. Future improvements and changes to the process configuration could alter these results.

### Appendix A-3—Sugar Separation

In this study, prehydrolyzed wood exits the prehydrolysis reactor with a high concentration of solids (approximately 24% solids with 10% xylose). At these conditions there is no free water, and all the water, xylose, acid, and other soluble materials are absorbed into the particle. The solids concentration of a completely saturated biomass particle is 18% to 20% (Schell 1990).



**Table A-3. Economy Summary of the Dilute Acid Process**

Grassroots plant, first-quarter 1990 cost

Plant Capacity: 160,000 dry lb/h

Total Capital Investment:

Aspen wood plant \$32.5 MM

Wheat straw plant \$34.3 MM

	Aspen	Wheat Straw
		¢/lb sugars
Raw Materials		
Feed	3.09	3.21
Acid	0.07	0.08
Lime	0.05	0.06
Utilities		
Process water	0.03	0.04
Cooling water	0.02	0.02
Steam-60 psig	0.21	0.15
Steam-600 psig	0.0	0.0
Electricity	0.35	0.09
Labor	0.06	0.06
Maintenance	0.10	0.11
Overhead	0.10	0.11
Insurance and taxes	0.05	0.06
By-product credits		
Lignin	1.00	0.90
Capital charges	0.75	0.82
Totals	3.88	3.91

Feedstock Cost:\$42/dry ton

FCR: 0.20

One technique for removing xylose, acid, and other solubles from the particles is by repeated washing with water, which allows the sugars to diffuse from the particles into the bulk solution. Further recovery can then be achieved by squeezing the particles (e.g., by centrifugation), thus extracting more liquid-containing sugars. The resulting liquid is then neutralized with lime, forming calcium sulfate (gypsum). After removal of the gypsum, the stream is sent to a xylose fermentation unit for conversion of xylose to ethanol. For this option, using two centrifuges in series with a counter current flow of wash water at an assumed rate of 2.5 lb wash water per pound of solids and dewatering to 35% solids, gave a 67% recovery of xylose and diluted the xylose stream from 10.0% to 6.7%. Sugar recovery can be improved by using more wash water, but the xylose is further diluted. In order to obtain a reasonable recovery of sugars (90%), it is estimated that four or five stages of centrifugation will be required at the same wash water flow rate. Using Bauger (1984) cost data for centrifuges doing similar service, the estimated purchase

price for five stages of centrifugation is \$8.0 MM and total capital cost of \$22.2 MM. This is roughly 16% of our estimated total capital cost of \$141 MM. With this option, additional capital is also required for gypsum separation equipment.

Because of the large capital expense associated with this option, it is preferable to neutralize the entire stream out of prehydrolysis and send the stream to xylose fermentation. During the fermentation, xylose will diffuse out of the particle into the surrounding liquid, where it will be converted into ethanol. Because xylose is disappearing from the bulk solution (via conversion to ethanol), a concentration gradient will exist that will continue to drive xylose from the particle into the surrounding liquid. The gypsum produced by neutralization and lignin will be carried along through xylose fermentation. The advantages of this option over the previous one are reduced cost (by elimination of centrifugation and gypsum separation) and potentially higher ethanol concentration from xylose fermentation. Because of these potential advantages, this option was used for this study. However, it is important to note that no actual performance data exist for this option.

#### **Appendix A-4—Cellulase Production**

Over the years, cellulase productivity has been improved through the development of new strains of the cellulase-producing fungus *Trichoderma reesei*. Researchers at Rutgers University developed a mutation of *T. reesei*, Rut C-30, which has significantly higher enzyme productivity than the previously used strain QM 9414. More recently, Cetus Corporation (Emeryville, CA) developed a highly productive mutation, L-27 (Shoemaker et al. 1981). In addition, experiments with various types of nutrient media have enabled cheaper ingredients, such as corn steep liquor, to be substituted for more expensive ingredients, such as proteose peptone.

Performance data for cellulase production using batch and fed-batch production techniques are available and listed in Table A-4. Fundamental kinetic information for the production of cellulase was not available. An analysis of the data shows that fed-batch production of cellulase has a higher productivity and yield than batch production. However, a true fed-batch mode of operation is not possible because the feed is wet. Fed-batch requires that essentially dry feed be added to the fermenter or that a portion of the fermenter contents be removed and slurried with the incoming feed. Because a fed-batch situation could not be envisioned, the process was designed for batch operation. Furthermore, the data are not of sufficient quality or quantity to allow any correlations to be developed. The values used for yield, substrate concentration, residence time, and specific activity are average values determined from the data for batch operation.

#### **Appendix A-5—Xylose Fermentation**

A preliminary analysis was performed to determine the relative merits of simultaneous fermentation and isomerization of xylose (SFIX) versus *E. coli* for the fermentation of xylose to ethanol. The genetically engineered *E. coli* supplied by L. Ingram has been shown to produce high ethanol yields of 88%-95% (Spindler 1989) but requires large amounts of base to neutralize fermentation acids. SFIX does not require large amounts of base but has lower yields and requires xylose isomerase production and immobilization. Both fermentations require 2 days to achieve adequate yields.

The purpose of this analysis is to determine the best option for xylose fermentation, SFIX or *E. coli*. Initially, we will assume that xylose isomerase is infinitely stable and 100% recoverable without cost; thus, there are no capital or operating expenses associated with isomerase use. Operating and capital costs, except for base addition, are assumed to be equal for both systems, since both fermentations require 2 days. Thus, under these conditions, if the extra revenues received for the increased ethanol production

Table A-4. Cellulase Production Data for *Trichoderma reesei*<sup>a</sup>

Cellulose Concentration (g/L)	Cell Density (g/L)	Specific Activity (IU/g eng.)	Residence Time (d)	Productivity <sup>a</sup> (IU/L-h)	Cellulase Yield (IU/g cellulose)	Culture <sup>f</sup> Method	Source
268 <sup>b</sup> (20) <sup>c</sup>	20	880	12	210	226	Fed-batch	Watson et al. 1984
150 <sup>b</sup> (20) <sup>c</sup>			12	105	203	Fed-batch	Hendy et al. 1982
100 <sup>b</sup>			12.5	87	262	Fed-batch	Hendy et al. 1984
80 <sup>b</sup> (20) <sup>a</sup>			10	159	477	Fed-batch	McLean et al. 1985
150 <sup>b</sup>			13	48	100	Fed-batch	Wilke, Blanch 1985
100 <sup>b</sup>			10	112	270	Batch	Wilke, Blanch 1985
50 <sup>a</sup>		1220	6	83	240	Batch	Sheir-ness et al. 1984
50 <sup>a</sup>		660	8	75	288	Batch	Tangnu et al. 1981
50 <sup>a</sup>			4	68	130	Batch	Hendy et al. 1982
70 <sup>a</sup>		800	5	167 <sup>e</sup>	286	Batch (150L)	Durand et al. 1988
50 <sup>a</sup>	11		7	71 <sup>e</sup>	240	Batch	Watson, Nelligan 1983
50 <sup>a</sup>			7	48	160	Batch	Schell et al. 1990
40 <sup>a</sup>	14	100	2.75	42	69	Batch	Mohagheghi et al. 1988

<sup>a</sup> Strain Rut C-30 except for Durand et al. (1988), CL-847; and Schell et al. (1990), L-27

<sup>b</sup> Total cellulose added divided by fermenter volume

<sup>c</sup> Initial cellulose concentration

<sup>d</sup> Maximum productivity except as noted, calculated from cellulase yield, cellulose concentration, and residence time

<sup>e</sup> Average productivity, maximum not available

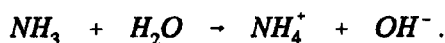
<sup>f</sup> Laboratory scale fermenters except as noted

Average productivity (fed-batch): 120      Average residence time (fed-batch): 11.6 days      Average residence time (batch): 5.7 days  
Average productivity (batch): 79      Average specific activity (all data): 732 IU/g  
Average cellulase yield (fed-batch): 256 IU/g      Average cellulase yield (batch): 202 IU/g

from the *E. coli* fermentation offset the base cost, this will be a more viable method for xylose fermentation.

D. Spindler's (1989) data for ethanol concentration (g/L) and NaOH usage (mL) as a function of fermentation pH are shown in Table A-5. Ethanol production from xylose fermentation (in gal) is calculated from the ethanol concentration and fermenter working volume of 2.0 L used in these fermentations. NaOH usage is calculated from the base molarity and amount used in milliliters. Because  $\text{NH}_4\text{OH}$  is a cheaper base and provides some nutritional requirements, the equivalent amount of  $\text{NH}_4\text{OH}$  that would be required is calculated from the ratio of molecular weights. The amount of  $\text{NH}_4\text{OH}$  required per gal of ethanol produced is the ratio of  $\text{NH}_4\text{OH}$  usage (in lb) to ethanol produced (gal). The base cost (\$/gal ethanol produced from xylose fermentation) is  $\text{NH}_4\text{OH}$  required (lb/gal ethanol) multiplied by  $\text{NH}_4\text{OH}$  cost developed below.

The cost of anhydrous ammonia is \$90/ton (*Chemical Marketing Reporter* 1989) (\$.045/lb). Ammonia dissociates into ammonium ions according to the following reaction:



One lb of ammonia produces 2.06 lb of ammonium hydroxide. The cost of ammonium hydroxide is then:

$$\frac{\$0.045}{\text{lb NH}_3} \left( \frac{1.0 \text{ lb NH}_3}{2.06 \text{ lb NH}_4\text{OH}} \right) = \frac{\$0.0218}{\text{lb NH}_4\text{OH}}.$$

**Table A-5. *E. coli* Fermentation Data and Base Usage and Cost**

pH	Ethanol		NaOH Usage <sup>b</sup>		NH <sub>4</sub> OH		
	Conc. (g/L)	Prod. <sup>a</sup> (gal)	(mL)	(g)	Usage (g)	Req/. (lb/gal)	NH <sub>4</sub> OH Cost (\$/gal)
7.0	36.6	.0248	250	50	43.7	3.89	0.84
6.5	39.3	.0266	240	48	42.0	3.48	.076
6.0	36.3	.0246	100	20	17.5	1.57	.034
5.5	22.6	.0153	380	76	66.5	9.57	.209

<sup>a</sup> 2.0 L working volume

<sup>b</sup> 5.0 M NaOH

The analysis continues by taking the ethanol yield for SFIX as 70% and the yield from the *E. coli* fermentation as 90%. (D. Spindler's yields for the *E. coli* fermentation were 90%, 96%, 89%, and 55% for pH controlled at 7.0, 6.5, 6.0, and 5.5, respectively.) Then, for 1 gal of ethanol produced by SFIX, 1.29 gal will be produced by the *E. coli* fermentation, and, with ethanol priced at \$0.60 per gal (the cost goal of the SERI/DOE alcohol fuels program), this gives extra revenue of \$0.17 for the *E. coli* fermentation. The additional cost to attain the extra revenue (at pH 7.0) is \$0.11 (\$0.084/gal × 1.29 gal).

Thus, the extra revenue exceeds the extra cost, and, even in light of the optimistic assumption regarding isomerase cost for the SFIX process, the *E. coli* fermentation appears to be the more economical process.

In reality, xylose isomerase is not very stable at the pHs used for xylose fermentation, as shown by Figures A-4 and A-5. These data are for the immobilization of *E. coli* xylose isomerase as reported by G. Means (1989) and coworkers at Ohio State University. The exception is the immobilization of enzyme on polyethyleneimine-glutaraldehyde-silica beads (PGS), which showed no degradation after 2 days. However, even with infinite enzyme life, recovery of the immobilized enzyme from the fermentation broth in the presence of lignocellulosic particles may be difficult.

Several other problems will increase the complexity of the SFIX process. *S. pombe* (the yeast used in SFIX) requires glucose for growth, which can either be bought or obtained from hydrolysis of cellulose. The cost of glucose (cost from *Chemical Marketing Reporter* Nov. 27, 1989) per gal of ethanol is shown in Figure A-6 as a function of time to discharge of the entire fermenter contents and the cell replacement rate per fermentation (assuming cell recycle). The fermenter must be periodically dumped as the level of nonviable cells builds up. This buildup as a function of cell replacement rate per fermentation is shown in Figure A-7. Thus, low cell replacement rates and long times between fermenter dumps would be necessary to achieve reasonable glucose cost and lower enzyme replacement rates. However, this would lead to problems with buildup of lignin and unconverted cellulose in the fermenter.

Another option is xylose-fermenting yeast, but these fermentations suffer from low yields (50%-70%) and typically have longer fermentation times (2-5 days) (Skoog and Hahn-Hagerdal 1988). Another option is to buy Novo Sweetzyme Q (cost from Don Krull 1989) instead of producing xylose isomerase. Figure A-8 shows the cost of enzyme/gal ethanol out of the plant as a function of time until the fermenter

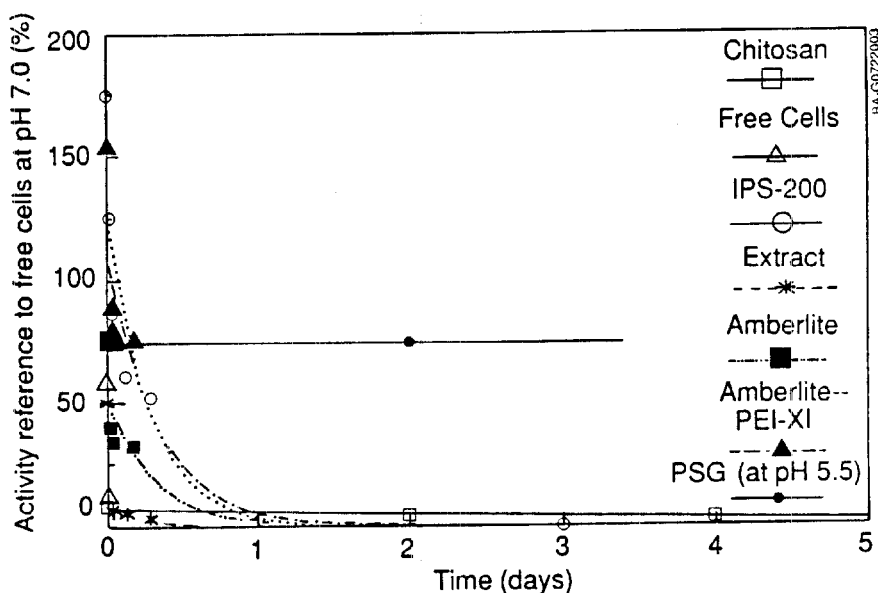


Figure A-4. Xylose isomerase stability at pH 5.75 for different immobilization methods

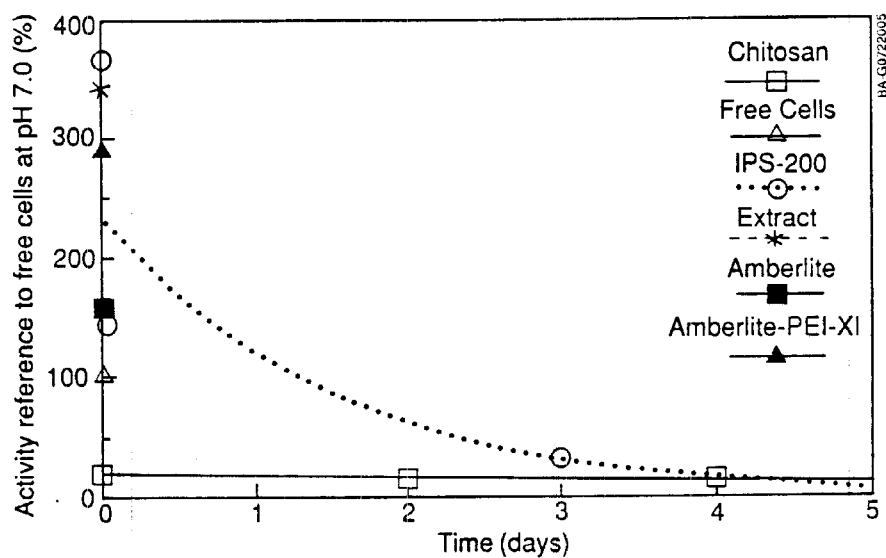


Figure A-5. Xylose isomerase stability at pH 6.0 for different immobilization methods

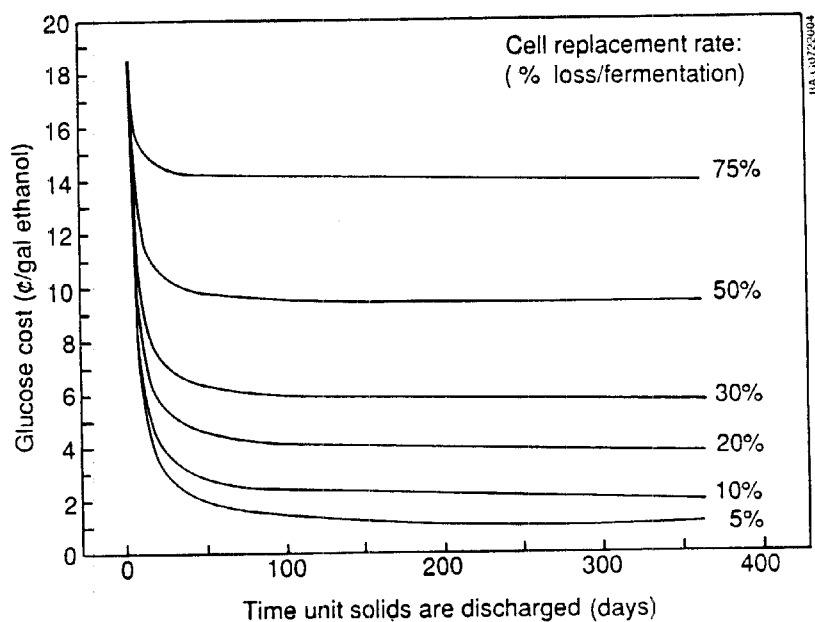


Figure A-6. Glucose cost as a function of discharge time and cell replacement rate

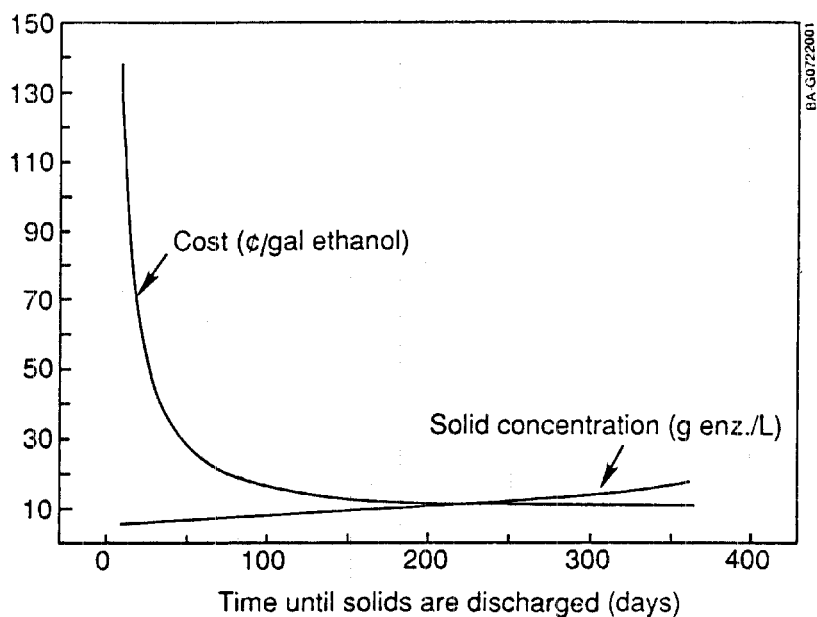


Figure A-7. Enzyme cost and concentration as a function of discharge time

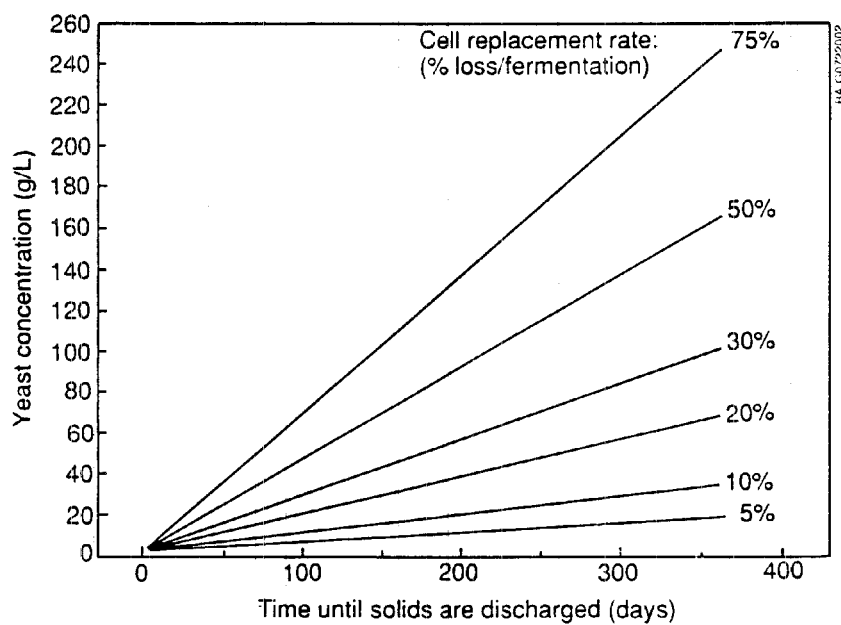


Figure A-8. Yeast concentration as a function of discharge time and cell replacement rate

is dumped. It is assumed that enzyme is added to maintain a constant enzyme loading and that the half-life of the enzyme is 220 days, which is estimated from Novo data. In fact, this half-life is probably much greater than the half-life that could be obtained at a pH of 5.75, which will make the true cost even higher. Even then, the minimum cost of the enzyme is approximately \$0.10/gal ethanol.

For reference, SFIX performance data have been compiled in Table A-6 and xylose isomerase production data are shown in Table A-7.

#### **Appendix A-6—Nutrient Cost for Xylose Fermentation and SSF**

Although this study assumed that all nutrient requirements are met by the recycle stream, a source of supplemental nutrients may be required. The cost for nutrients is calculated for the worst-case scenario, assuming that the recycled water does not contribute any nutrients. The nutrient requirements for both xylose fermentation and SSF are given in Table A-8. The media for xylose fermentation is an M9 minimal media with the following changes:  $\text{Na}_2\text{HPO}_4$  is eliminated because buffering is not required; the concentration of  $\text{KH}_2\text{PO}_4$  is doubled to supply additional phosphate; and it is assumed that only one amino acid is required at a concentration of 0.075 g/L.

The cost for nutrients is shown in Table A-9 for both xylose fermentation and SSF. The second column is the average cost for all nutrients and is determined by weighing the individual nutrient cost with the required concentration. If nutrients are required at the concentrations assumed in Table A-8, then the cost of nutrients for xylose fermentation and SSF is 21¢/gal and 14¢/gal of denatured fuel, respectively, for a total cost of 35¢/gal of denatured fuel. This is not an insignificant cost, thus, the nutrient requirements for the process are extremely important.



Table A-6. Simultaneous Isomerization and Fermentation Data

Xylose Concentration (g/L)	Isomerase Loading (IU/g)	Temp. Residence Time (°C)	Ethanol Yield (%)	Cell Density (g/L)	Organism	Culture Method	Source
60	4.2 <sup>a</sup>	35	71		<i>S. pombe</i>	Batch	Lastick et al. 1989
60	8.3 <sup>b</sup>	28	50	2.5	<i>S. pombe</i>	Batch	Orton et al. 1988
120	10.4 <sup>b</sup>	35	63	45	<i>S. cerevisiae</i>	Batch	Chiang et al. 1981
127		30	49	75	<i>S. cerevisiae</i>	Fed-	Hahn-Hagerdal et al. 1986
120	20.8 <sup>b</sup>	40	85		<i>S. cerevisiae</i>	Batch	Gong et al. 1981
50		29	9	1.1	<i>S. pombe</i>	Batch	Wang et al. 1980

<sup>a</sup> Optimum rate at 2.3 IU/g

<sup>b</sup> Assumes 25 IU/g enzyme for Sweetzyme Q (Lastick et al. 1986)

Table A-7. Xylose Isomerase Production Data

Substrate	Substrate Concentration (g/L)	Specific Activity (IU/mg)	Residence Time (h)	Isomerase Yield (IU/g)	Cell Density (g/L)	Organism	Source
Glucose		1.2	4			<i>E. coli</i>	Lastic et al. 1986
Glucose	35.5 <sup>a</sup>		14 <sup>b</sup>	40	15	<i>E. coli</i>	Spencer 1989
Glucose	1.0	.286				<i>E. coli</i>	Stavis, Ho 1985
Xylose	2.0	.560				<i>E. coli</i>	Woycha et al. 1983
Glucose		.015				<i>E. coli</i>	Schellenburg et al. 1983
Xylose		.179				<i>E. coli</i>	Schellenburg et al. 1983
Xylose/Glycerol	20/20	.259				<i>E. coli</i>	Schellenburg et al. 1983
Xylose	30.0	.260				<i>S. violaceus</i>	Callens et al. 1985

<sup>a</sup> Fed-batch production

<sup>b</sup> First 12 h at 32°C, then temperature raised to 42°C for last 2 h.

**Table A-8. Concentration and Cost for Xylose Fermentation and SSF Nutrients**

Nutrient	Concentration (g/L)	Cost (¢/lb)
<b>Xylose Fermentation</b>		
CaCl <sub>2</sub>	0.01	7.65
MgSO <sub>4</sub>	0.12	14.00
KH <sub>2</sub> PO <sub>4</sub>	7.00	6.60
NaCl	3.50	1.00
NH <sub>4</sub> Cl	1.00	18.00
Amino acid	0.075	1000.00
<b>SSF</b>		
(NH) <sub>2</sub> SO <sub>4</sub>	1.50	4.25
MgSO <sub>4</sub>	0.10	14.00
CaCl <sub>2</sub>	0.06	7.65
Corn steep liquor	7.50	11.00

Source for SSF nutrients: University of Arkansas

**Table A-9. Nutrient Cost**

	Average Nutrient Cost (¢/lb)	Annual Cost (\$MM)	Cost (¢/gal fuel)
Xylose Fermentation	12.35	12.24	21.1
<b>SSF</b>			
Nutrients	3.96	0.61	1.1
Corn steep liquor	11.00	7.60	13.1
Total SSF Cost	14.96	8.21	14.2
Total Cost		20.45	35.2

## Appendix B

### Process Technical Data and Assumptions

#### General Specifications

Mixing motors	1.0 hp/1,000 gal except as noted
Tank capacity	80% full except as noted
Chilled water temperature	10°C
Steam levels	50 psig, 150 psig

#### Feedstock Composition (dry):

Cellulose	46.2%
Xylan	24.0%
Lignin	24.0%
Ash	0.2%
Others (soluble)	5.6%

The wood is delivered to the plant at 50% moisture.

#### Feed Handling

##### Wood Chip Pile:

Storage	4 days
Losses	none (assumption)

##### Mill:

Particle size	2.0-3.0 mm
Power requirement	128 hp-h/dry ton (vendor number)

#### Pretreatment

Pretreatment is a two-step process. Acid impregnation is followed by prehydrolysis at a higher temperature.

##### Impregnation:

Reactor	continuous digester (Carpenter 20 alloy)
Temperature	100°C
Pressure	atmospheric
Exit solids concentration	35 wt % (design assumption)
Residence time	10 min. (assumption)

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### Prehydrolysis:

Reactor	continuous digester (Carpenter 20 alloy)
Temperature	160°C
Pressure	105 psig
Acid concentration	0.85 wt % after steam addition in the prehydrolysis reactor
Residence time	10 min

### Conversions:

Cellulose to glucose	3.0% (kinetic data)
Cellulose to HMF	0.1% "
Unconverted cellulose	96.9% "
Xylan to xylose	80.0% (experimental data)
Xylan to furfural	13.0% "
Unconverted xylan	7.0% "

### Flash Tank:

Solids concentration	12.0 wt % (Assumed to be a pumpable slurry)
Residence time	5 min (design assumption)
Pressure	atmospheric
Mixing power	2.0 hp/1,000 gal (assumption)
All furfural to overheads	(assumption)

### Neutralization

Neutralizing agent	lime
Residence time	10 min (assumption)
Mixing power	2.0 hp/1,000 gal (assumption)

### Cellulase Production

#### Fermenters:

Type	batch
Temperature	28°C
Pressure	10.0 psig (design assumption)
pH	4.8
Neutralizing agent	NH <sub>3</sub>
Ammonia usage	0.045 lb/lb cellulose and xylose (experimental data)
Ammonia tank size	1% of fermentation capacity (assumption)
Substrate	Cellulose and xylose
Substrate concentration	5.0% (conforms to most experimental data)
Fermentation time	5.5 days (experimental data)
Cycle time	6.0 days "
Cellulase yield	202 IU/g cellulose and xylose (experimental data)
Enzyme activity	732 IU/g enzyme (experimental data)
Final cell density	20 g/L (experimental data)
Specific growth rate	0.0415/h "
O <sub>2</sub> uptake Rate	42 mM O <sub>2</sub> /L-h "
Dissolved O <sub>2</sub>	20% of air saturation (design assumptions)
Antifoam use	1.0 ml/L of fermenter volume (assumption)

Nutrients:	(from literature reference)
Ammonium sulfate	1.4 g/L
Potassium phosphate	2.0 g/L
Magnesium sulfate*7H <sub>2</sub> O	0.3 g/L
Calcium chloride*2H <sub>2</sub> O	0.4 g/L
Tween 80	0.2 g/L
Corn steep liquor	15.0 g/L

Seed Fermenters (same as above except):

Fermentation time	3.5 days (from literature reference)
Cycle time	4.0 days "
Pressure	atmospheric
Substrate concentration	1.0% cellulose and xylose
Cell yield	0.5 g cells/g substrate (assumption)
Final seed volume	5.0% of fermenter volume (assumption)
Airflow	0.2 vvm (assumption)
Mixing power	
First seed vessel	0.5 hp/1,000 gal (assumption)
Other seed vessels	1.0 hp/1,000 gal "

Sterile Feed Tank:

Mixing power	maximum - 2.0 hp/1,000 gal (assumption)
	average - one-half maximum

Cellulase Hold Tank:

Mixing power	maximum - same as cellulase fermenters
	average - one-half maximum

## Xylose Fermentation

Fermenters:

Type	continuous stirred tanks in series
Temperature	37°C
Pressure	atmospheric
pH	7.0
Neutralizing agent	NH <sub>3</sub>
Ammonia use	0.2878 lb/lb ethanol produced (experimental data)
Ethanol yield	85.5% (experimental data and assumption of 90% recovery of xylose from the particles)
Mixing power	0.1 hp/1,000 gal (assumption) \
Tank fill	95%
Fermentation time	2 days (experimental data)
Nutrients	none required (assumed contained in recycle water)
Ethanol to vent	Aspen simulation (82% recovery)

Seed Fermenters (same as above except):

Type	batch
Fermentation time	12 h
Cycle time	1 day
Substrate	xylose and glucose
Substrate concentration	2.0%
Cell yield	0.5 g cells/g xylose (assumption)
Final seed volume	10.0% of fermenter volume (design assumption)
Airflow	0.2 vvm (assumption)

Seed Hold Tank:

Mixing power	maximum - 0.1 hp/1,000 gal (assumption) average - one-half maximum
--------------	---

**Cellulose Fermentation**

Fermenters:

Type	continuous stirred tanks in series, 32 tanks
Temperature	37°C
Pressure	atmospheric
Fermentation time	7 days (experimental data)
Tank fill	95%
pH	Uncontrolled
Cell yield	0.5 g cells/g cellulose (assumption)
Mixing power	0.1 hp/1,000 gal (assumption)

Conversions:

Cellulose to ethanol	72.0% (experimental data)
Cellulose to fusel oils	0.1% (assumption)
Cellulose to glycerol/ acetaldehyde	4.9% (assumption)
Cellulose to cells	10.0% "
Xylan to xylose	80.0% "
Nutrients	none required (assumed contained in recycle water)
Enzyme loading	7 IU/g cellulose
Ethanol tolerance	4.5%
Ethanol to vent	Aspen simulation (82% recovery)

Seed Fermenters:

Same as above except

Type	batch
Fermentation time	
<i>S. cerevisiae</i>	1 day
<i>B. clausenii</i>	2 day
Cycle time	
<i>S. cerevisiae</i>	1.5 days
<i>B. clausenii</i>	2.5 days

Substrate - Initial	glucose
- Final seed vessel	cellulose/cellulase
Substrate concentration	1.0%
Final seed volume	10% of fermenter volume for each culture (design assumption)
Airflow	0.2 vvm (assumption)
Mixing power	
First seed vessel	0.5 hp/1,000 gal (assumption)
Other seed vessels	1.0 hp/1,000 gal (assumption)

#### Seed Hold Tanks:

Mixing power	maximum -0.1 hp/1,000 gal (assumption) average - one-half maximum
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### Ethanol Purification

#### Distillation:

Ethanol concentration (Rectification column)	95.0 wt %
Water to fusel oils	5.0 lb/lb fusel oils (Badger data)
Ethanol concentration (Beer column)	40.0 wt % (Badger data)
Reflux ratio (beer)	0.4 (Badger data)
Reflux ratio (rectification)	1.6 (Badger data)

#### Lignin Separation:

Solids recovery (centrifugation)	95% (assumption)
Solid concentration (centrifugation)	50% "

### Wastewater Treatment

#### Anaerobic Digestion:

Organics converted	90% (lignin unconverted) (experimental data, Rivard 1990)
Biogas production	0.8 lb gas/lb organics converted, balance to cell mass (experimental data, Rivard 1990)

#### Aerobic Digestion:

All remaining organics degraded except lignin	
Solids recovery (centrifugation)	100% (assumption)
Solid concentration (centrifugation)	50% "



## Utilities

Boiler includes Flakt drying system:

Design pressure	1100 psig, 300°F superheat
Efficiency	83.5%

Turbogenerator:

Reduces 1100 psig steam to 150 psig and 50 psig for process use, any remaining steam is condensed	
Efficiency	78.5%

Boiler Feed Water System:

Water rate to boiler	3.0% of steam usage plus direct injection (assumption)
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Cooling Water System:

Water losses	1.3% of flow for evaporation (from literature)
	0.3% of flow for windage "
	2.7% of flow for blowdown "

Sterile Air System:

Air temperature	28°C
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## Appendix C

### Process Data

#### Appendix C-1—Heat Capacities

Wood	0.32 Btu/lb-°F (Wenzel 1970)
Sulfuric acid	0.37 Btu/lb-°F (Himmelblau 1962)
Ethanol vapor	0.40 Btu/lb-°F (Yaws 1977)
Lime (solid)	0.29 Btu/lb-°F (Himmelbeau 1974)
Gypsum	0.26 Btu/lb-°F (Touloukian and Buyco 1970)
Air	0.25 Btu/lb-°F (McCabe and Smith 1976)
Corn steep liquor	1.00 Btu/lb-°F (assumed)
Ethanol	0.35 Btu/lb-°F (Touloukian and Buyco 1970)
Carbon dioxide	0.21 Btu/lb-°F (McCabe and Smith 1976)
Corn oil	0.51 Btu/lb-°F (Perry and Chilton 1973)
Water vapor	0.45 Btu/lb-°F (Touloukian and Buyco 1970)
Water	1.00 Btu/lb-°F

#### Appendix C-2—Densities

All process streams were estimated as water	62.4 lb/ft <sup>3</sup>
Ethanol	48.7 lb/ft <sup>3</sup> (Weast 1972)
Sulfuric acid	114.2 lb/ft <sup>3</sup> (Weast 1972)
Lime	139.8 lb/ft <sup>3</sup> (Weast 1972)
Corn oil	57.4 lb/ft <sup>3</sup> (Weast 1972)

### Appendix C-3—Higher Heating Values

Lignin	11478 Btu/lb (Shafizadeh 1984)
Cellulose	7464 Btu/lb (Shafizadeh 1984)
Methane	23984 Btu/lb (Himmelblau 1974)
Ethanol	12836 Btu/lb (Weast 1972)
Xylose	6747 Btu/lb (Weast 1972)
Xylan	7464 Btu/lb (assumed the same as cellulose)
Soluble solids	5000 Btu/lb (assumed)
Cellulase	5000 Btu/lb (assumed)
Glycerol	7774 Btu/lb (Weast 1972)
Acetaldehyde	12835 Btu/lb (Himmelblau 1974)
Methane	23984 Btu/lb (Himmelblau 1974)

### Appendix C-4—Latent Heat

Steam (50 psig)	912 Btu/lb (Steam tables)
Steam (150 psig)	857 Btu/lb (Steam tables)
Ethanol (12°C)	423 Btu/lb (Touloukian and Buyco 1970)
Ethanol (100°C)	324 Btu/lb (Touloukian and Buyco 1970)

### Appendix C-5—Heat Transfer Coefficients (Tubular Exchangers)

Condensing steam-liquid	700 Btu/°F ft <sup>2</sup> -h (Perry and Chilton 1973)
Liquid-liquid	225 Btu/°F ft <sup>2</sup> -h (Perry and Chilton 1973)
Condensing vapor-gas	100 Btu/°F ft <sup>2</sup> -h (Perry and Chilton 1973)
Condensing vapor-liquid	400 Btu/°F ft <sup>2</sup> -h (assumed)
Coils (coils in agitated tank)	100 Btu/°F ft <sup>2</sup> -h (Perry and Chilton 1973)
Gas-liquid	60 Btu/°F ft <sup>2</sup> -h (Perry and Chilton 1973)

## Appendix C-6—Solubilities

Gypsum

0.222 g/100 cc (100°C) (Weast 1972)

0.241 g/100 cc (20°C) (Weast 1972)

## **Appendix D**

### **Spreadsheet Model**

A spreadsheet model of the biomass-to-ethanol process was developed the help perform sensitivity analyses on the conceptual process design. The model includes a complete material and energy balance, capital and operating cost estimates, and economic evaluation.

The material and energy balance includes 80 streams with up to 27 components, 6 utility summaries, and 14 chemical requirement summaries. There are approximately 100 process variables that may be manipulated in the material balance in order to carry out sensitivity analyses. The variables are listed in Table D-1.

The utility summaries generated by the material and energy balance include the following:

- Electricity
- Low-pressure steam
- High-pressure steam
- Cooling water
- Chilled water
- Fermentation air

Feedstock, catalysts, and chemicals summaries generated include the following:

- Biomass
- H<sub>2</sub>SO<sub>4</sub>
- Lime
- NH<sub>3</sub>
- Corn steep liquor
- Nutrients
- Antifoam
- Glucose
- Gasoline
- Diesel
- Makeup water
- Solids disposal
- BFW chemicals
- Cooling water chemicals

The capital cost estimate is generated using capacity exponents and a base case design for which a detailed cost estimate was originally made. The plant is broken down into 17 process areas and 9 utility areas as follows:

- Wood Handling
- Prehydrolysis
- Xylose Fermentation
  - Seed fermenters
  - Main fermenters
  - Remaining equipment
- Cellulase Production
  - Seed fermenters
  - Main fermenters
  - Remaining equipment
- Simultaneous Saccharification and Fermentation
  - Seed fermenters, culture 1
  - Seed fermenters, culture 2
  - Main fermenters
  - Remaining equipment
- Ethanol Recovery
  - Rectification column
  - Remaining equipment
- Off-site Tankage
- Environmental Systems
  - Wastewater treatment
  - Vent system
- Utilities
  - BFW, steam, and condensate
  - Boiler
  - Process water
  - Turbogenerator
  - Cooling water
  - Chilled water
  - Fermentation air
  - Auxilliary utilities

Economic analyses are done on a total-plant basis and a process-unit allocated-cost basis. The cost of ethanol production is determined on both a per-year and per-gallon basis. There are approximately 25 variables in the economic analysis section of the model. These variables include:

- Capital Cost
  - Exponents for cost of scaled equipment, by area
  - Installation factors, by area
  - Working capital
- Operating Costs
  - Onstream factor
  - Unit costs for all feedstocks, chemicals and utilities
  - Labor costs

**Table D-1. Variables in the Spreadsheet Model of the Biomass-to-Ethanol Process**

---

**Wood Handling**

Biomass feed rate, lb/h (wet basis)  
 Biomass composition, wt %, (wet basis)  
 Feed temperature, °F

**Prehydrolysis**

H<sub>2</sub>SO<sub>4</sub> feed rate, lb/h  
 H<sub>2</sub>SO<sub>4</sub> temperature, °F  
 Water temperature, °F  
 Low-pressure steam latent heat, Btu/lb  
 High-pressure steam latent heat, Btu/lb  
 Prehydrolysis reactor conversions  
     Xylan to xylose, %  
     Xylan to furfural, %  
     Xylan unconverted, %  
     Cellulose to glucose, %  
     Cellulose to hydroxymethylfurfural, %  
     Cellulose unconverted, %  
 Latent heat of stream from blowdown, Btu/lb  
 Dilution water rate to blowdown tank, lb/h

**Cellulase Production**

Fraction of hydrolyzate to cellulase production  
 Fraction to seed fermenters  
 Dilution water rate to seed fermenters, lb/h  
 Dilution water rate to main fermenters, lb/h  
 Cell mass production ratio in seed fermenters, lb/lb cellulose + xylose  
 Nutrient feed rate, g/L  
 Corn steep liquor rate, g/L  
 Base feed rate, lb NH<sub>3</sub>/lb cellulose + xylose  
 Antifoam feed rate, mL/L  
 Enzyme yield, IU/g cellulose + xylose  
 Enzyme specific activity, IU/g enzyme  
 Cell mass production ratio in main fermenters, lb/lb cellulose + xylose  
 Fermentation time, days  
 Fermentation air rate, vvm  
 Number of seed trains operating  
 Time between seed batches, h  
 Seed fermentation air rate, vvm  
 Agitator power for main fermenters, hp/1,000 gal  
 Agitator power for seed fermenters, hp/1,000 gal

**Table D-1. Variables in the Spreadsheet Model of the Biomass-to-Ethanol Process (Continued)**

**Xylose Fermentation**

Fraction of remaining hydrolyzate to seed fermenters  
 Dilution water rate to seed fermenters, lb/h  
 Cell mass production ratio in seed fermenters, lb/lb glucose + xylose  
 Base feed rate, lb  $\text{NH}_3$ /lb ethanol produced  
 Fermentation time, days  
 Fraction of xylose available for conversion  
 Fraction of available xylose converted to ethanol  
 Fraction of glucose converted to ethanol  
 Water in fermentation off gas  
 Ethanol in fermentation off gas  
 Number of seed trains operating  
 Time between seed batches, h  
 Seed fermentation air rate, vvm  
 Agitator power for main fermenters, hp/1,000 gal  
 Agitator power for seed fermenters, hp/1,000 gal

**Simultaneous Saccharification and Fermentation (SSF)**

Fraction to seed fermenters  
 Percent cellulose converted in seed fermenters  
 Percent glucose converted in seed fermenters  
 Cell mass production ratio in seed fermenters, lb/lb glucose  
 Dilution water rate to SSF, lb/h  
 SSF fermenter conversions  
     Cellulose to ethanol +  $\text{CO}_2$ , %  
     Cellulose to acetaldehyde + glycerol +  $\text{CO}_2$ , %  
     Cellulose to yeast +  $\text{CO}_2$ , %  
     Cellulose to fusel oils, %  
     Glucose conversion to products above, %  
     Xylan to xylose, %  
 Fermentation time, days  
 Water in fermentation off gas  
 Ethanol in fermentation off gas  
 Fraction of total water condensed from off gas  
 Fraction of total ethanol condensed from off gas  
 Number of seed trains operating  
 Split between two seed trains if two organisms used  
 Time between seed batches, h  
 Seed fermentation air rate, vvm  
 Agitator power for main fermenters, hp/1,000 gal  
 Agitator power for seed fermenters, hp/1,000 gal



**Table D-1. Variables in the Spreadsheet Model of the Biomass-to-Ethanol Process (Concluded)**

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**Ethanol Recovery**

- Percent of ethanol recovered
- Percent of acetaldehyde vented
- Percent of fusel oil recovered
- Weight percent fusel oil in fusel oil product (balance assumed to be water)
- Water to fusel oil decanter, number times fusel oil rate
- Weight percent ethanol in product from ethanol distillation
- Gasoline addition rate, lb gasoline/lb ethanol
- Fraction solids recovered in lignin centrifugation
- Weight percent solids in underflow from lignin centrifuge
- Recycled process water rate, lb/h

**Environmental systems**

- Wastewater from CIP/CS, lb/h
- Fraction of organics converted in anaerobic digestion
- Organics to biogas production ratio, lb/lb

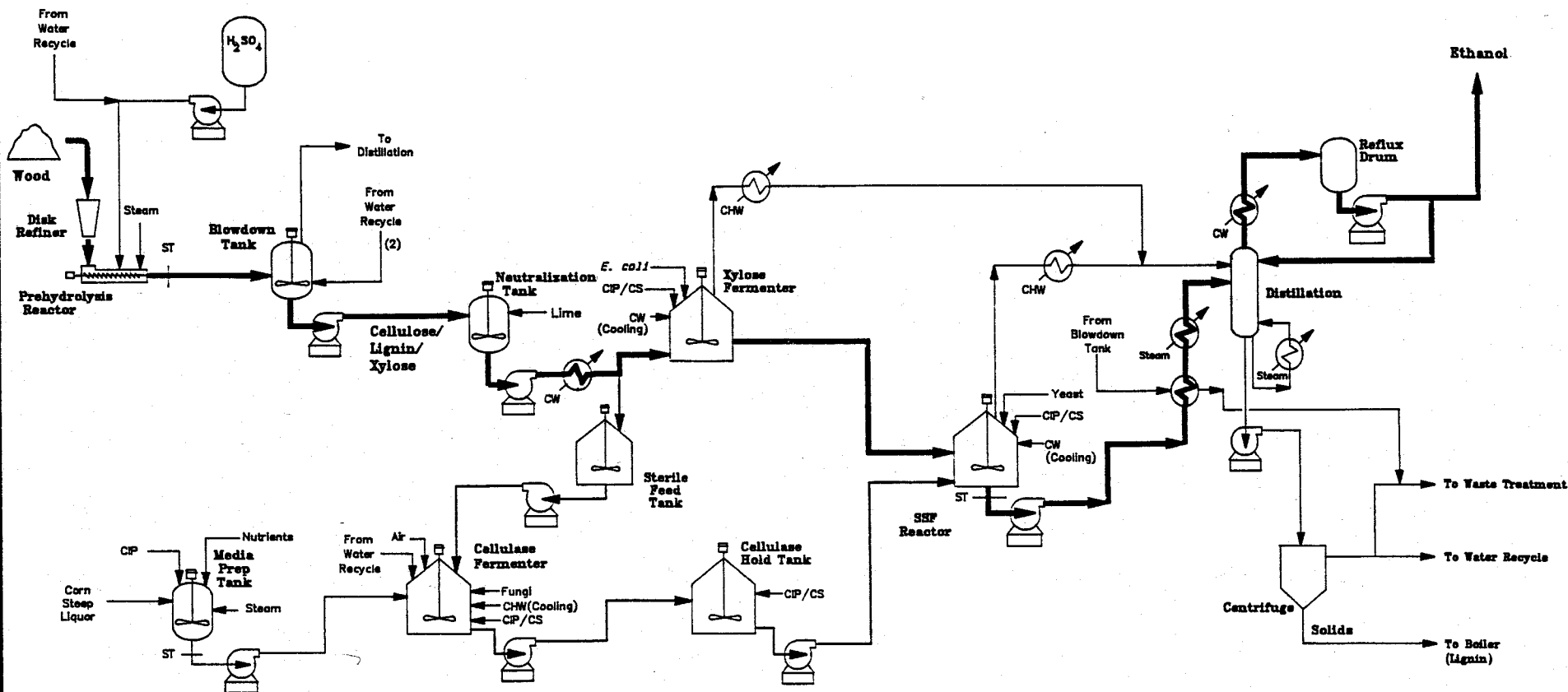
**Utilities**

- Boiler efficiency, %
- Turbogenerator efficiency, %

## **Appendix E**


### **Process Flow Diagrams**

This section contains the process flow diagrams with material balances for the currently designed biomass-to-ethanol plant. Also included is a plot plan for the entire plant and a more detailed plot plan for the fermentation area.

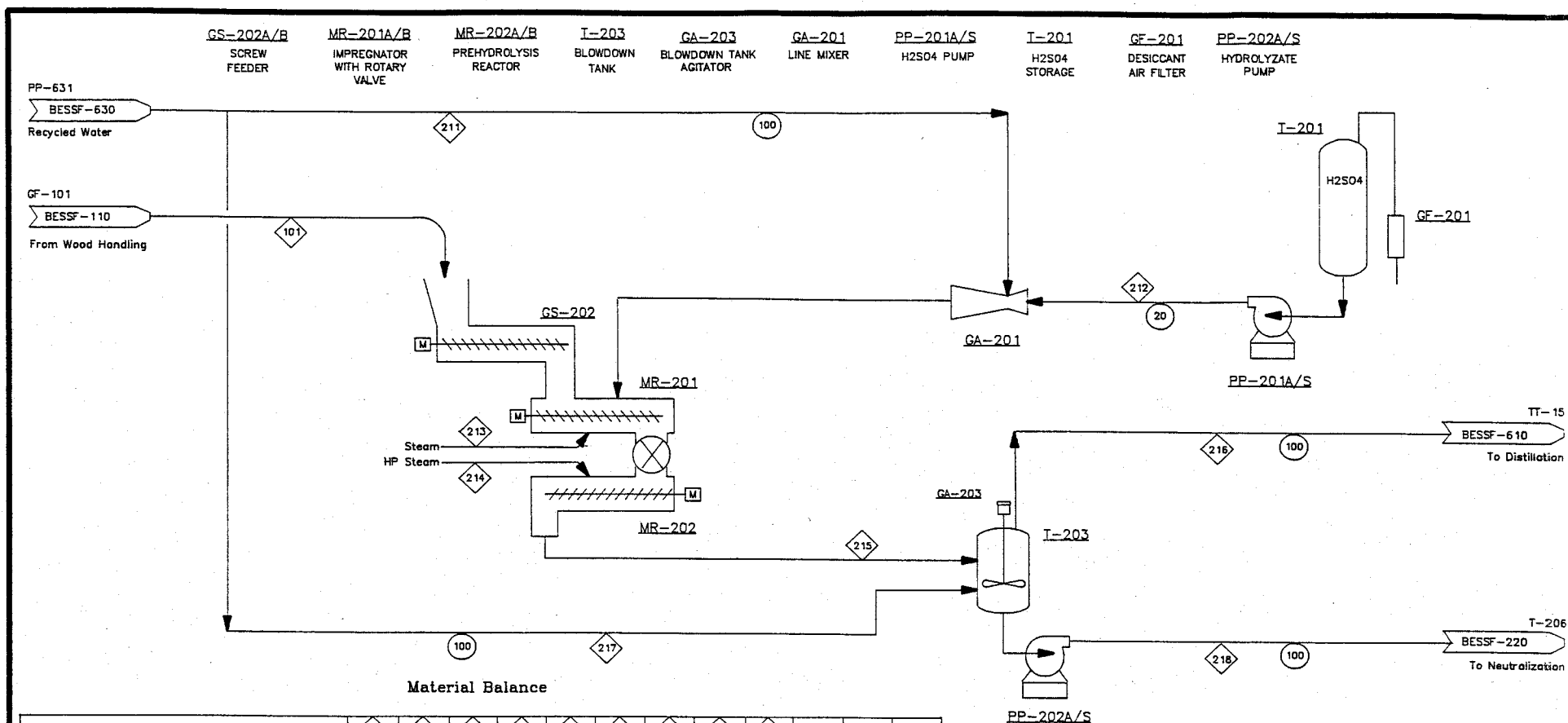


# Notes

- (1) ST: Sterile Boundary; CIP/CS: Clean in Place/Chemical Sterilization; CW: Cooling Water; CHW: Chilled Water
- (2) Dilution water to be added if substrate concentration is such that the resulting ethanol concentration in xylose fermentation or SSF would exceed the respective ethanol tolerances or to achieve a pumpable wood slurry

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Project No DOE/APP/BESSF-10/89				
Draftsperson P. WALTER	Date:			
Designer	Date:			
Approval N. HINMAN	Date:			
Revision Code		SERI Drawing No.: BESSF-022		Sheet 1 of 1
Misc. Data		Ref. Dwg. No.		Date 6/25/90





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SIMULTANEOUS SACCHARIFICATION & FERMENTATION  
PREHYDROLYSIS

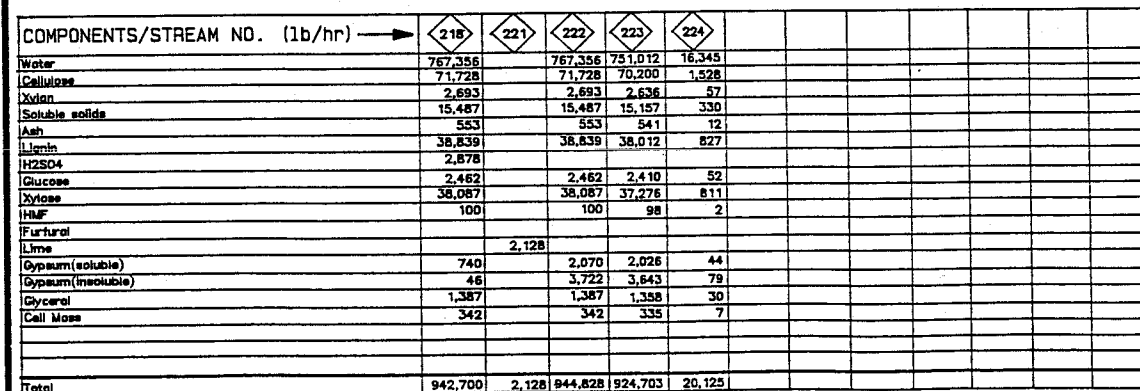
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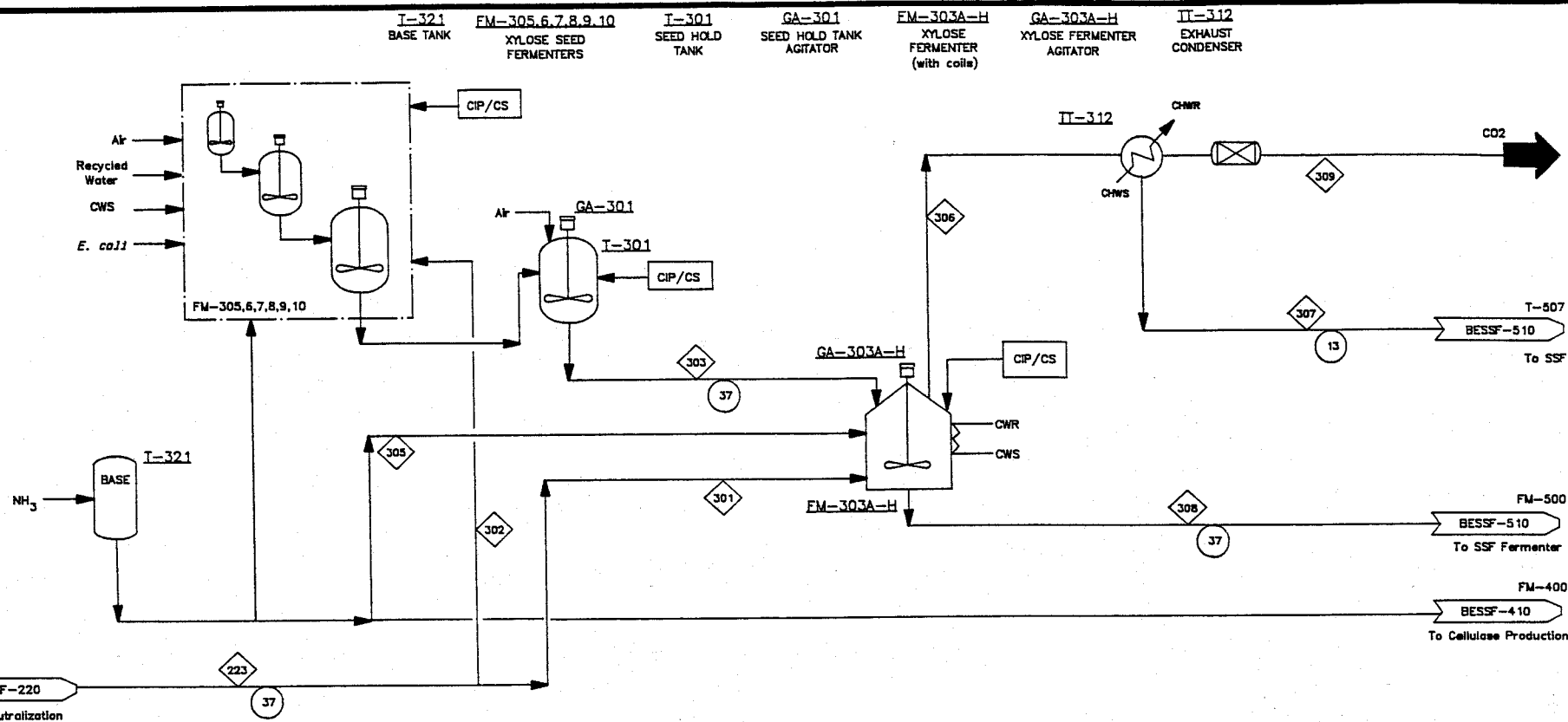
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Torget et al.(1988)

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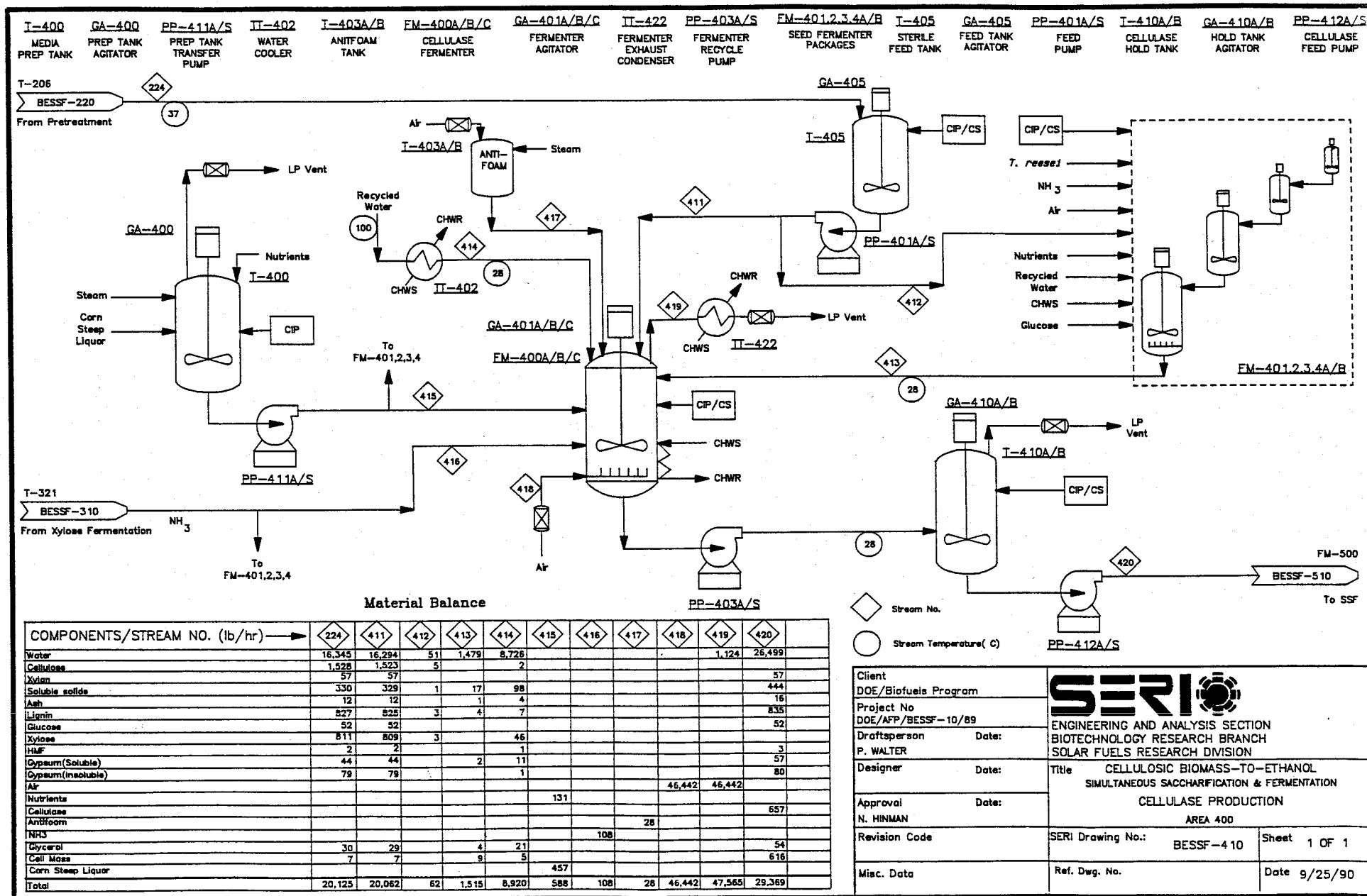
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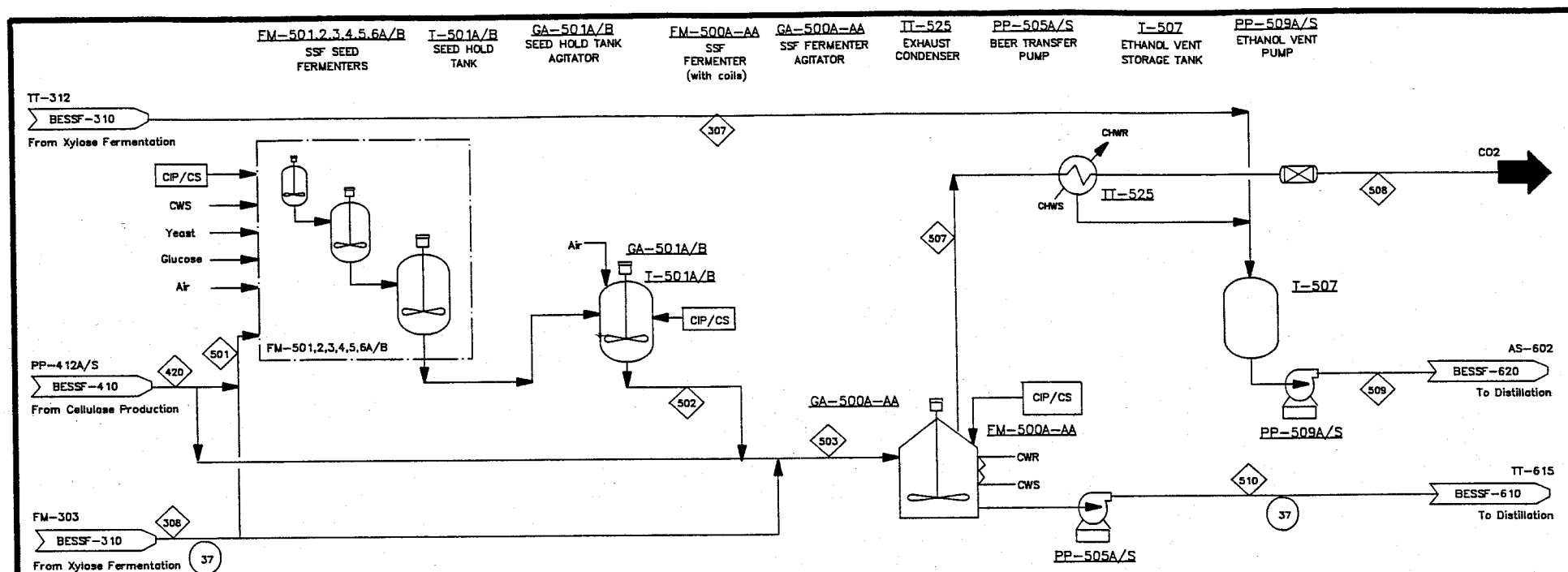
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		Sheet 2 OF 2  Date 9/25/90



Client DOE/Biofuels Program		<b>SERI</b> ENGINEERING AND ANALYSIS SECTION BIOTECHNOLOGY RESEARCH BRANCH SOLAR FUELS RESEARCH DIVISION
Project No. DOE/AFB/BESSF-10/89		
Draftsperson P. WALTER	Date:	Title CELLULOSIC BIOMASS-TO-ETHANOL SIMULTANEOUS SACCHARIFICATION & FERMENTATION XYLOSE FERMENTATION AREA 300
Designer	Date:	
Approval N. HINMAN	Date:	
Revision Code	SERI Drawing No.: BESSF-310	Sheet 1 of 1
Misc. Data	Ref. Dwg. No.	Date 9/25/90







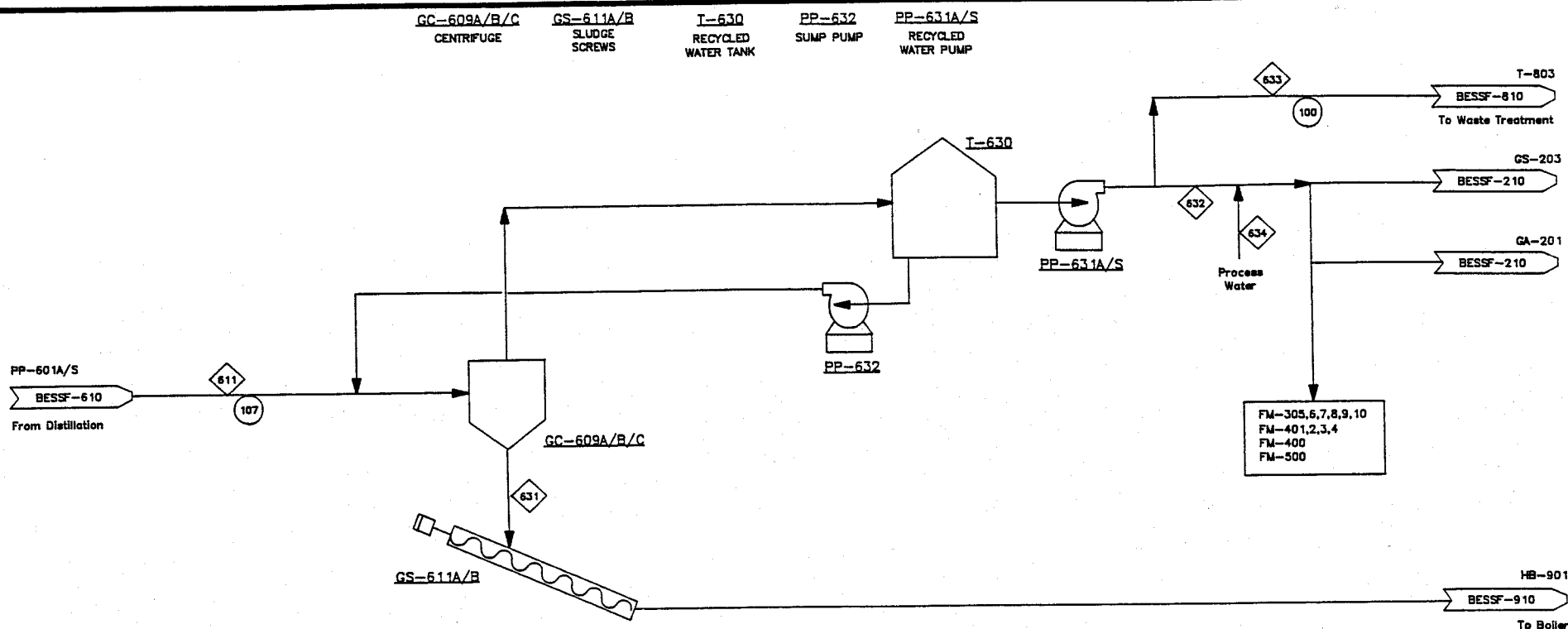
Material Balance

COMPONENTS/STREAM NO. (lb/hr) →	307	308	420	501	502	503	507	508	509	510		
Water	243	809,549	26,499	167,209	167,209	836,047	655	163	735	832,639		
Cellulose		70,210		14,042	11,991	68,159				8,861		
Xylan		2,637	57	539	539	2,694				539		
Soluble solids		15,819	444	3,253	3,253	16,263				16,263		
Ash		565	16	116	116	581				581		
Lignin		38,057	835	7,778	7,778	38,892				38,892		
Glucose		335	52	77	310							
Xylose		5,181		1,036	1,036	5,181				7,638		
H <sub>2</sub> O		102	3	21	21	105				105		
Gypsum(soluble)		2,101	57	432	432	2,158				2,158		
Gypsum(insoluble)		3,648	80	746	746	3,728				3,728		
CO <sub>2</sub>							27,457	27,068				
Cellulose			657	131	131	657				657		
Ethanol	908	15,333		3,067	3,067	15,333	4,237	621	4,525	39,007		
Fusel Oils										76		
Glycerol		1,498	54	310	310	1,592				3,456		
Acetaldehyde										911		
Cell Mass		1,346	616	392	1,570	3,139				6,940		
Total	1,151	966,380	29,369	199,150	198,198	994,798	32,349	27,652	5,259	962,499		

Client DOE/Biofuels Program	<b>SERI</b>		
Project No DOE/AFB/BESSF-10/89	ENGINEERING AND ANALYSIS SECTION		
Draftsperson P. WALTER	Date:	BIOTECHNOLOGY RESEARCH BRANCH	
Designer	Date:	SOLAR FUELS RESEARCH DIVISION	
Approval N. HINMAN	Date:	Title CELLULOSIC BIOMASS-TO-ETHANOL	
Revision Code		SIMULTANEOUS SACCHARIFICATION & FERMENTATION	
Misc. Data		AREA 500	
		Ref. Dwg. No.	Date
		BESSF-510	9/25/90






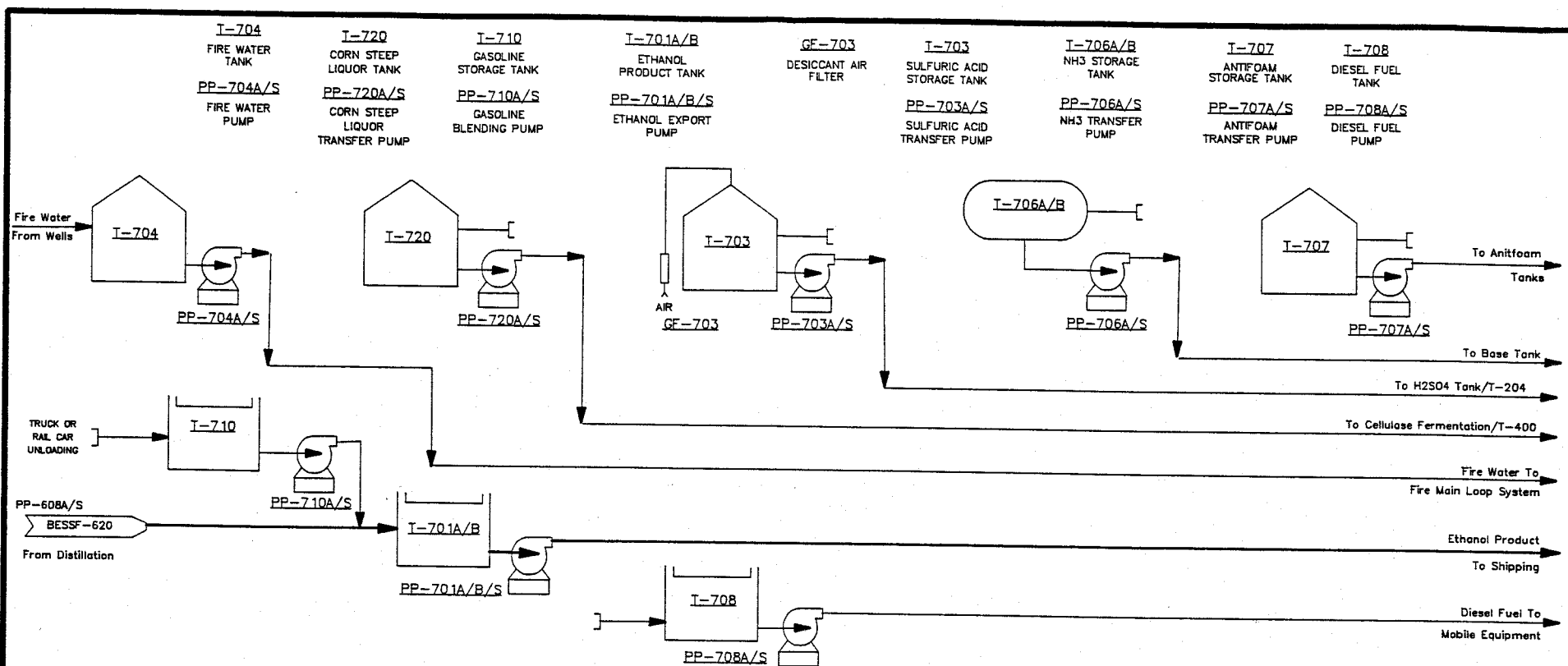



### Material Balance

COMPONENTS/STREAM NO.	(lb/hr)	$\diamond$ 811	$\diamond$ 831	$\diamond$ 832	$\diamond$ 833	$\diamond$ 834
Water		831,428	55,671	373,525	402,229	276,628
Cellulose		8,861	8,628	112	121	
Xylan		539	525	7	7	
Soluble solids		16,267	1,089	7,308	7,869	
Ash		581	39	281	281	
Lignin		38,892	37,871	492	529	
Xylose		7,640	512	3,432	3,696	
HMF		103	7	47	51	
Gypsum(soluble)		1,846	124	829	893	
Gypsum(insoluble)		4,041	3,933	51	55	
Cellulose		657	44	295	318	
Glycerol		3,457	231	1,553	1,672	
Cell Moss		6,940	6,738	88	94	
Total		921,249	115,433	388,000	417,816	276,628

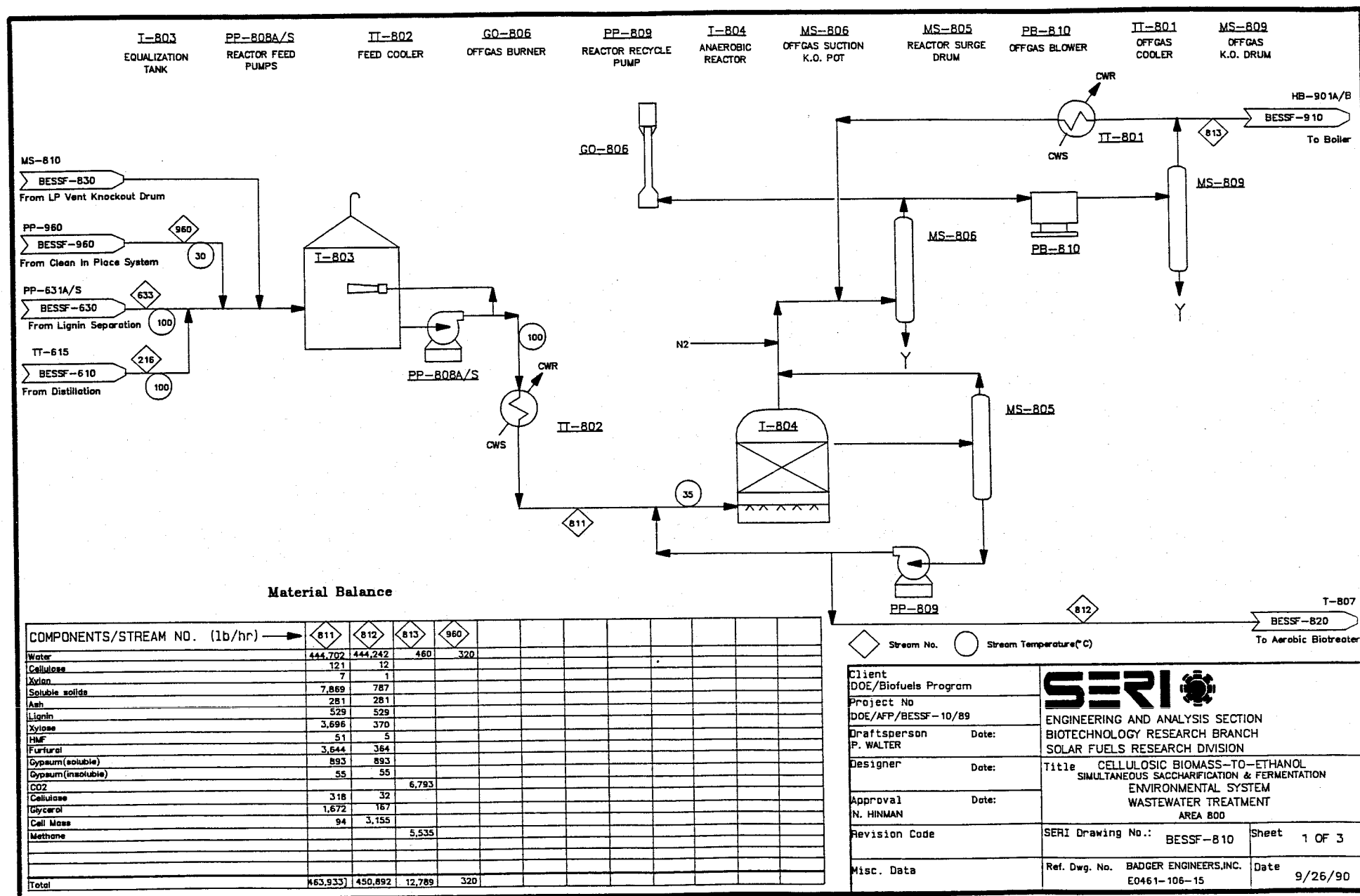
 Stream No.      Stream Temperature(°C)

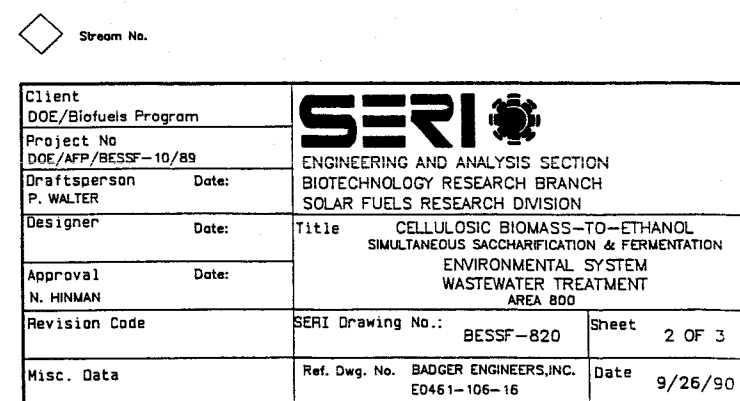
Client DOE/Biofuels Program	<div>SERI</div> <div>ENGINEERING AND ANALYSIS SECTION BIOTECHNOLOGY RESEARCH BRANCH SOLAR FUELS RESEARCH DIVISION</div>		
Project No DOE/AFB/BESSF-10/89			
Draftsperson P. WALTER			Date:
Designer			Date:
Approval N. HINMAN			Date:
		Title CELLULOSIC BIOMASS-TO-ETHANOL SIMULTANEOUS SACCHARIFICATION & FERMENTATION	
		LIGNIN SEPARATION AREA 600	
Revision Code	SERI Drawing No.: BESSF-630	Sheet 3 OF 3	
Misc. Data	Ref. Dwg. No.	Date 9/25/90	



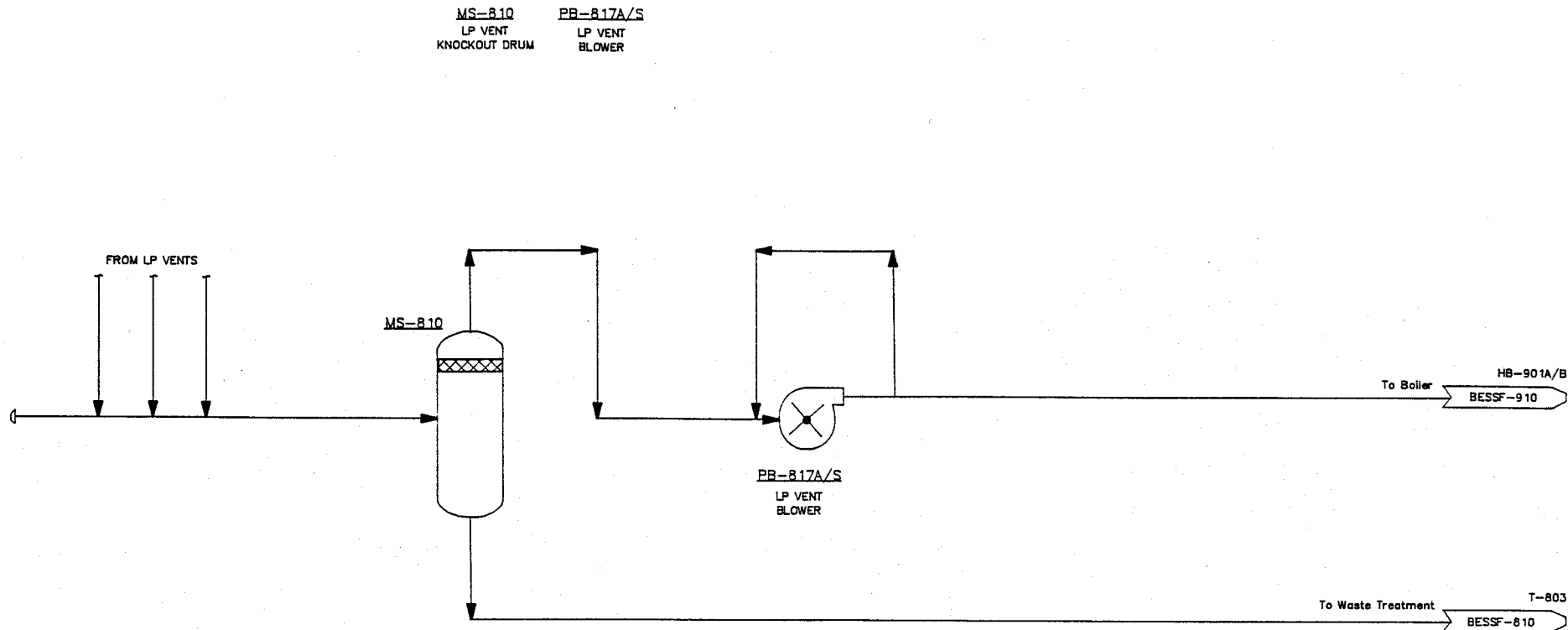
Client DOE/Biofuels Program		 ENGINEERING AND ANALYSIS SECTION BIOTECHNOLOGY RESEARCH BRANCH SOLAR FUELS RESEARCH DIVISION	
Project No DOE/AFP/BESSF-10/89			
Draftsperson P. Walter	Date:		
Designer	Date:		
Approval N. Hinman	Date:		
Revision Code			
Misc. Data		Ref. Dwg. No. BADGER ENGINEERS, INC. ED461-106-13	Date 9/25/90


SERI Proprietary Information  
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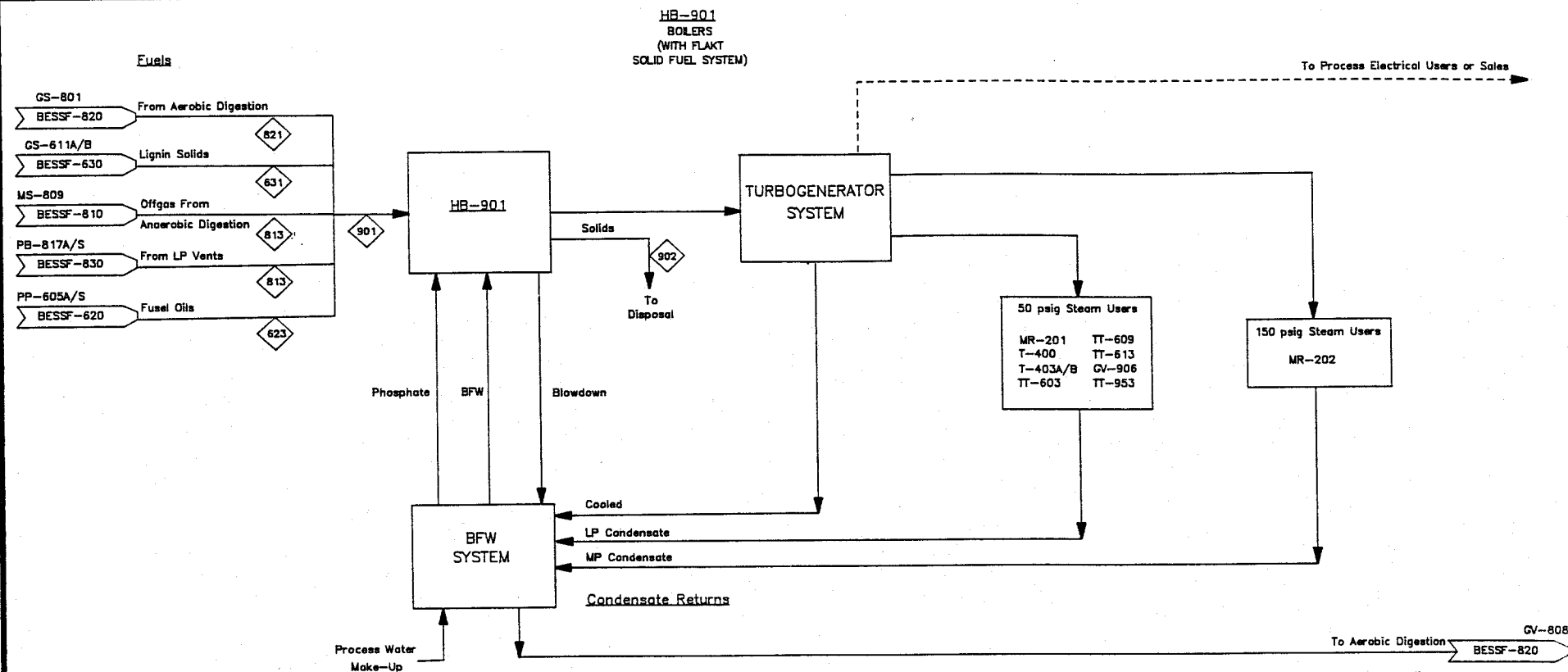
COMPONENTS/STREAM NO. (lb/hr) →	812	821	822
Water	444,242	4,577	439,665
Cellulose	12		
Xylan	1		
Soluble solids	787		
Ash	281		281
Lignin	529	529	
Xylose	370		
HMF	5		
Furfural	364		
Gypsum(Insoluble)	893	893	
Gypsum(soluble)	55		
Cellulose	32		
Glycerol	167		
Cell Mass	3,155	3,155	
Total	450,892	9,154	439,946



Client DOE/Biofuels Program	<b>SERI</b>  ENGINEERING AND ANALYSIS SECTION BIOTECHNOLOGY RESEARCH BRANCH SOLAR FUELS RESEARCH DIVISION	
Project No DOE/AFP/BESSF-10/89		
Draftsperson P. WALTER	Date:	Title CELLULOSIC BIOMASS-TO-ETHANOL SIMULTANEOUS SACCHARIFICATION & FERMENTATION ENVIRONMENTAL SYSTEM PLANT VENT SYSTEM AREA 800
Designer	Date:	
Approval N. HINMAN	Date:	
Revision Code	SERI Drawing No.: BESSF-830	Sheet 3 OF 3
Misc. Data	Ref. Dwg. No. BADGER ENGINEERS, INC. E0461-106-17	Date 6/21/90


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### Material Balance

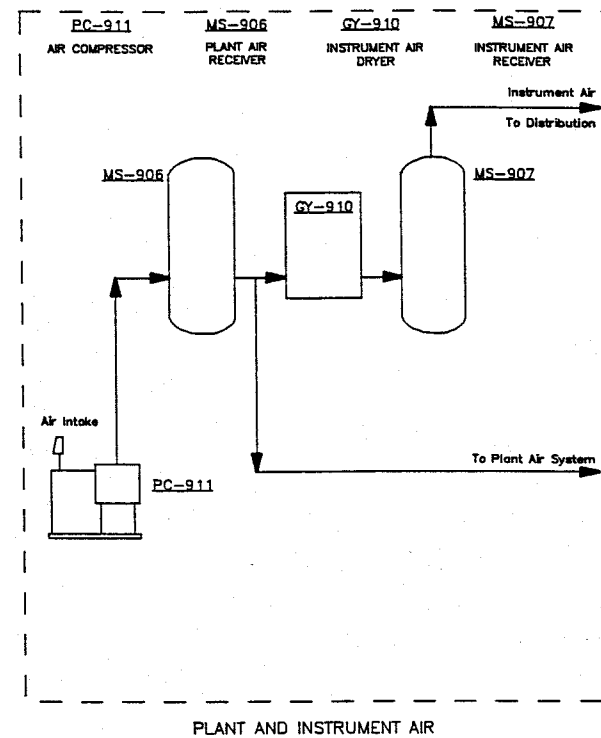
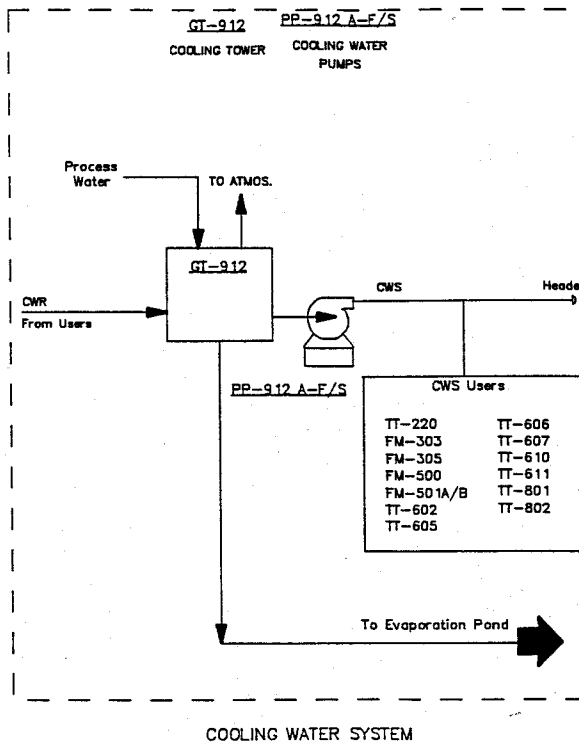
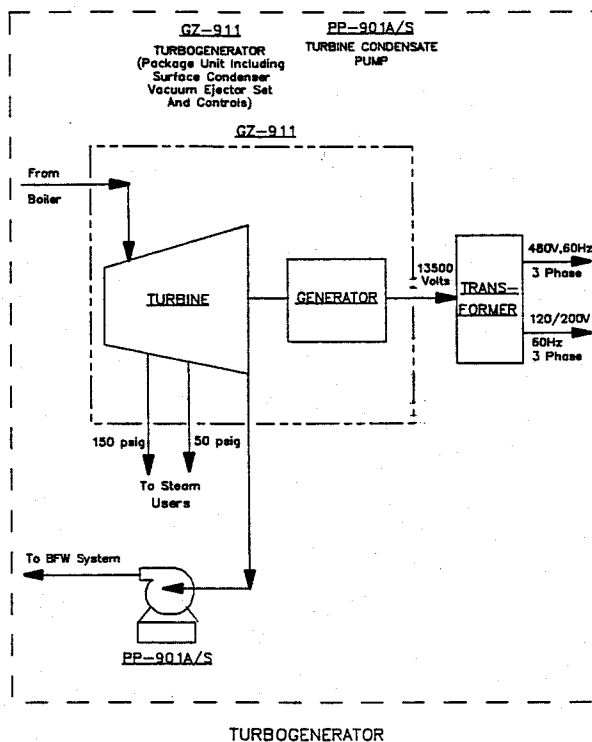
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 Stream No.

Client DOE/Biofuels Program				
Project No DOE/AFP/BESSF-10/89			ENGINEERING AND ANALYSIS SECTION	
Draftsperson P. WALTER	Date:		BIOTECHNOLOGY RESEARCH BRANCH	
Designer	Date:		SOLAR FUELS RESEARCH DIVISION	
Approval N. HINMAN	Date:		CELLULOSIC BIOMASS-TO-ETHANOL SIMULTANEOUS SACCHARIFICATION & FERMENTATION UTILITIES BOILERS & STEAM DISTRIBUTION AREA 900	
Revision Code		SERI Drawing No.: BESSF-910		Sheet 1 OF 6
Misc. Data		Ref. Dwg. No. BADGER ENGINEERS, INC. E0461-105-18		Date 9/26/90







Client DOE/Biofuels Program		 <b>ENGINEERING AND ANALYSIS SECTION</b> <b>BIOTECHNOLOGY RESEARCH BRANCH</b> <b>SOLAR FUELS RESEARCH DIVISION</b>  <b>CELLULOSIC BIOMASS-TO-ETHANOL</b> <b>SIMULTANEOUS SACCHARIFICATION &amp; FERMENTATION</b>  <b>UTILITIES</b> <b>AREA 900</b>
Project No DOE/AFP/BESSF-10/89		
Draftsperson P. WALTER	Date:	
Designer	Date:	
Approval N.HINMAN	Date:	
Revision Code	SERI Drawing No.:  BESSF-940	
Misc. Data	Ref. Dwg. No. BADGER ENGINEERS, INC. E0461-106-21,22,23	Sheet 4 OF 6
		Date 9/26/90

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PK-950A/B/S  
AIR COMPRESSOR  
PACKAGE

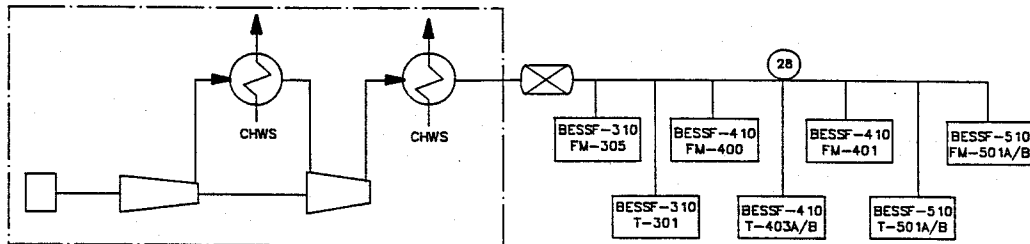
PK-951  
CHILLED WATER  
PACKAGE

TT-953  
WATER  
STERILIZER

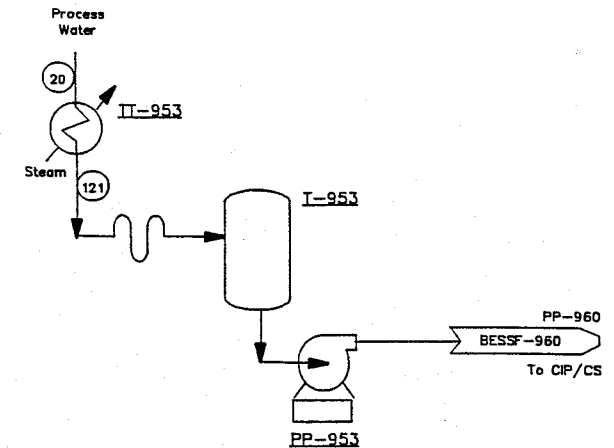
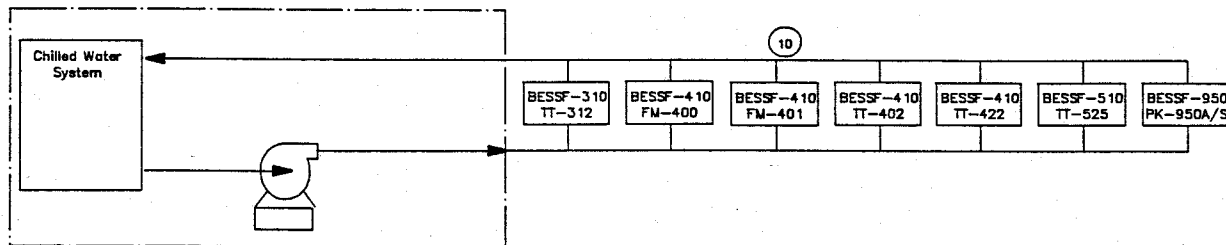
T-953  
STERILE  
WATER TANK

PP-953  
STERILE  
WATER PUMP


PK-950A/B/S



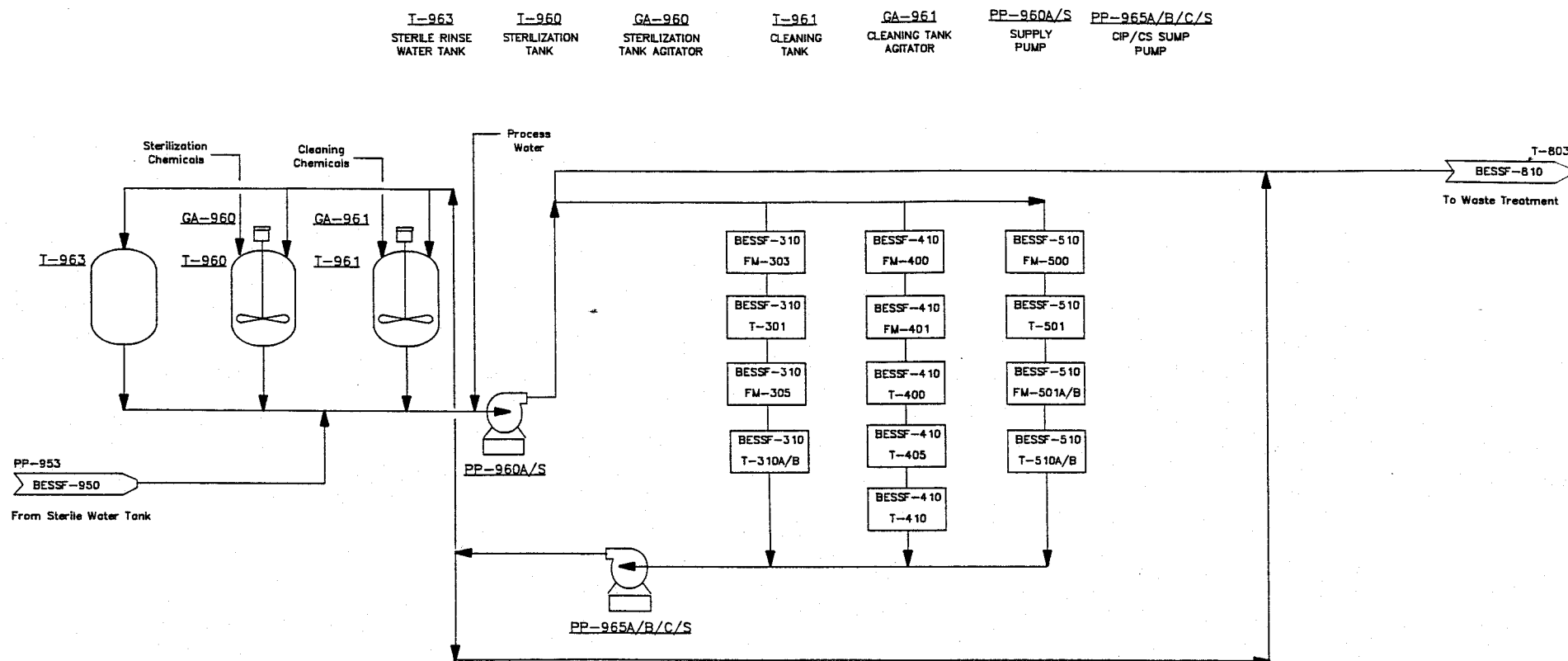
PK-951




○ Stream Temperature(°C)

Client DOE/Biofuels Program		 ENGINEERING AND ANALYSIS SECTION BIOTECHNOLOGY RESEARCH BRANCH SOLAR FUELS RESEARCH DIVISION		
Project No DOE/AFP/BESSF-10/89				
Draftsperson P. WALTER	Date:			
Designer	Date:			
Approval N. HINMAN	Date:			
Revision Code		SERI Drawing No.: BESSF-950		Sheet 5 of 6
Misc. Data		Ref. Dwg. No.		Date 9/26/90

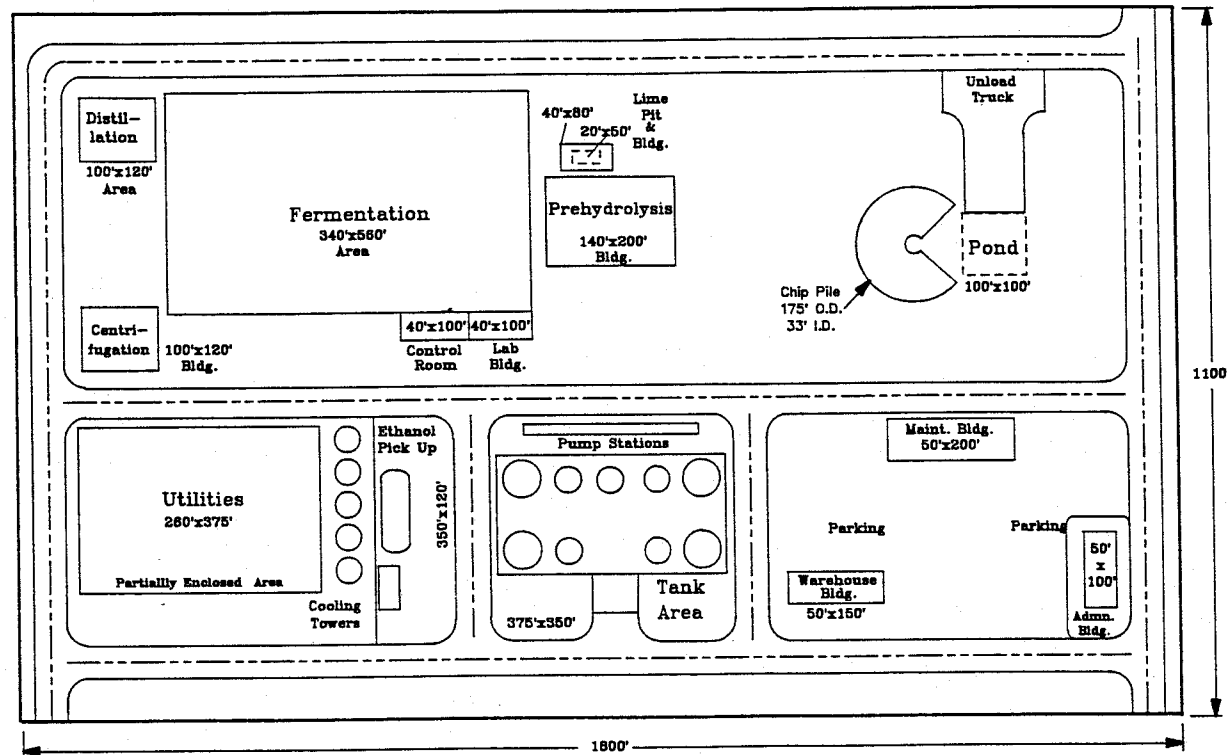
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Client DOE/Biofuels Program		 ENGINEERING AND ANALYSIS SECTION BIOTECHNOLOGY RESEARCH BRANCH SOLAR FUELS RESEARCH DIVISION	
Project No DOE/AFB/BESSF-10/89			
Draftsperson P. WALTER	Date:		
Designer	Date:		
Approval N. HINMAN	Date:		
Revision Code		SERI Drawing No.: BESSF-960	Sheet 6 of 6
Misc. Data		Ref. Dwg. No.	Date 9/26/90

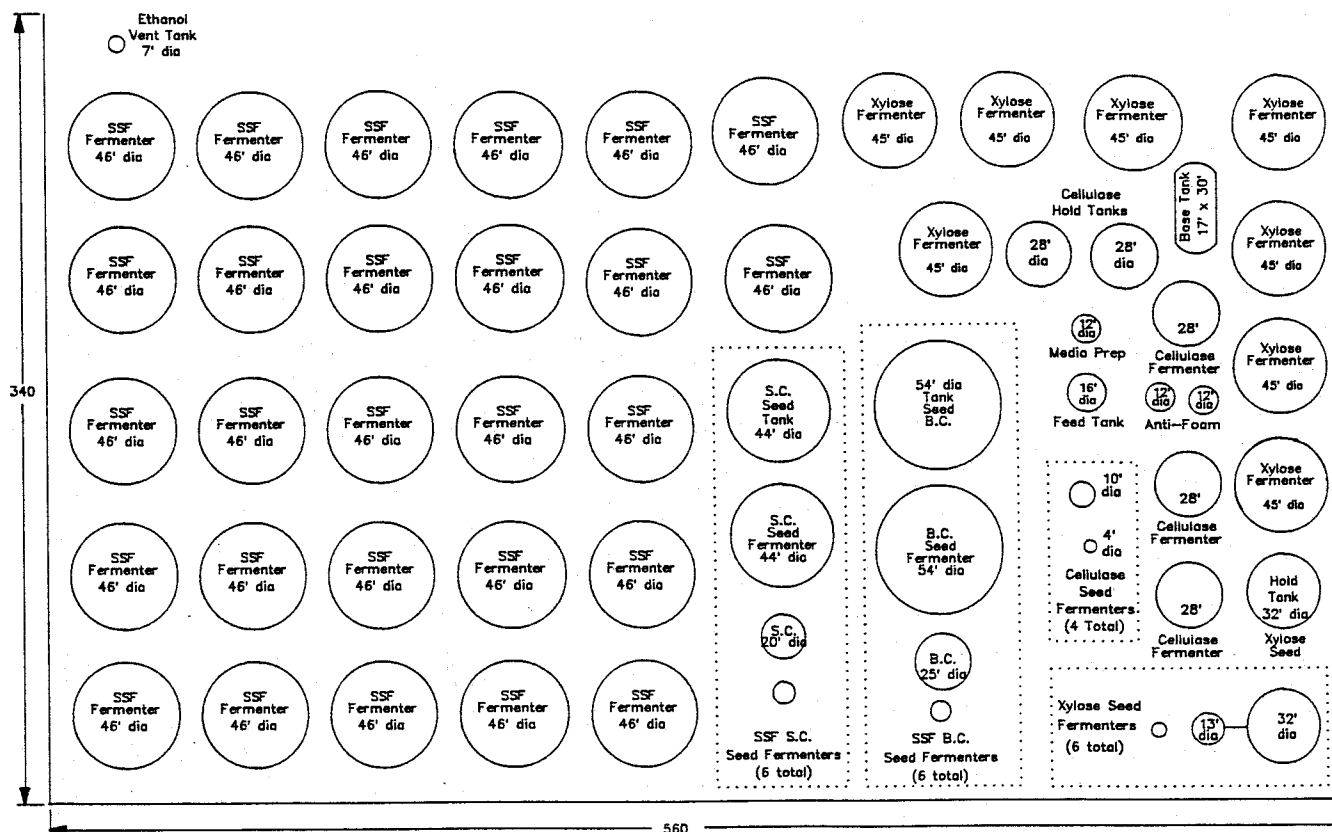
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# Ethanol Plant Model



Client DOE/Biofuels Program		 ENGINEERING AND ANALYSIS BIOTECHNOLOGY RESEARCH BRANCH SOLAR FUELS RESEARCH DIVISION	
Project No DOE/AFP/BESSF-10/89			
Draftsperson P. WALTER	Date:		
Designer P. WALTER	Date:	Title CELLULOSIC BIOMASS-TO-ETHANOL SIMULTANEOUS SACCHARIFICATION & FERMENTATION  Plot Plan	
Approval N. HIRMAN D. SCHILL	Date:		
Revision Code		SERI Drawing No.: BESSF-1000	Sheet
Misc. Data 1" = 200'		Ref. Dwg. No.	Date 8/28/90

SERI Proprietary Information  
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Client DOE/Biofuels Program		 ENGINEERING AND ANALYSIS BIOTECHNOLOGY RESEARCH BRANCH SOLAR FUELS RESEARCH DIVISION	
Project No DOE/AFB/BESSF-10/89			
Draftsperson P. WALTER	Date:		
Designer	Date:		
Approval N. HINMAN	Date:		
Revision Code		Title CELLULOSIC BIOMASS-TO-ETHANOL SIMULTANEOUS SACCHARIFICATION & FERMENTATION FERMENTATION TANK LAYOUT	
Misc. Data		Ref. Dwg. No.	Sheet 1 OF 1
		Date 8/27/90	

SERI Proprietary Information  
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Biomass To Ethanol  
Simultaneous Saccharification & Fermentation  
Project No. DOE/AFR/BEESF/02

## EQUIPMENT LIST

ITEM #	EQUIPMENT NAME	ACTUAL FRACTION FULL	DIA/HEIGHT	PRESS.	TEMP.	H	MAT'L	Capacity gal	DRW #	NO. REQ'D	PURCHASED COST/UNIT	TOTAL PURCHASED COST	SOURCE
T-201	Sulfuric Acid Storage	0.770	14ft/11ft	Atm	100 F	V	fiberglass	5800	BEESF-210	1	\$6,500	\$6,500	Hall et al. 1995
T-202	Blowdown Tank	0.499		Atm	100 V	V	SS 304	18000	BEESF-210	1	\$50,000	\$50,000	Icarus 1985
T-206	Neutralization Reaction Tank	0.781	22ft/22ft	50	250 V	V	SS 304	24000	BEESF-220	1	\$60,000	\$60,000	Icarus 1985
FM-302A-H	Xylose Fermenter	0.944	45ft/50ft	Atm	150 V	V	CS	746000	BEESF-310	8	\$190,000	\$1,520,000	Icarus 1985
FM-305	Xylose Seed Fermenter	0.935	32ft/50ft	50	150 V	V	CS	300000	BEESF-310	1	\$88,000	\$88,000	Icarus 1985
FM-306	Xylose Seed Fermenter	0.941	14ft/25ft	50	150 V	V	CS	29900	BEESF-310	1	\$25,000	\$25,000	Icarus 1985
FM-307	Xylose Seed Fermenter	0.946	6.5ft/12ft	50	150 V	V	CS	2990	BEESF-310	1	\$8,000	\$8,000	Icarus 1985
FM-308	Xylose Seed Fermenter	0.958	3ft/5ft	50	150 V	V	CS	310	BEESF-310	1	\$3,000	\$3,000	Ulrich 1984
FM-309	Xylose Seed Fermenter	0.952	1.5ft/2.2ft	50	150 V	V	CS	33	BEESF-310	1	\$1,500	\$1,500	Ulrich 1984
FM-310	Xylose Seed Fermenter	0.712	7ft/11ft	50	150 V	V	CS	1	BEESF-310	1	\$700	\$700	Ulrich 1984
T-301	Seed Hold Tank	0.935	32ft/50ft	10	150 V	V	CS	300000	BEESF-310	1	\$65,000	\$65,000	Icarus 1985
T-321	Base Tank	0.869	17ft/30ft	250	100 H	H	CS	13000	BEESF-310	1	\$25,000	\$25,000	Icarus 1985
FM-400A/B/C	Cellulase Fermenter	0.791	46ft/50ft	15	250 V	V	CS	214000	BEESF-410	3	\$65,000	\$195,000	Icarus 1985
FM-401A/B	Cellulase Seed Fermenter	0.759	15ft/30ft	50	250 V	V	CS	11300	BEESF-410	2	\$20,300	\$40,600	Icarus 1985
FM-402A/B	Cellulase Seed Fermenter	0.735	6ft/10ft	50	250 V	V	CS	550	BEESF-410	2	\$3,300	\$6,600	Icarus 1985
FM-403A/B	Cellulase Seed Fermenter	0.801	2ft/4ft	50	250 V	V	CS	25	BEESF-410	2	\$1,100	\$2,200	Icarus 1985
FM-404A/B	Cellulase Seed Fermenter	0.721	7/2ft	50	250 V	V	CS	1.4	BEESF-410	2	\$600	\$1,200	Icarus 1985
T-400	Media Prep Tank	0.773	12ft/18ft	50	250 V	V	CS	9400	BEESF-410	1	\$15,000	\$15,000	Icarus 1985
T-403A/B	Antifoam Tank	0.828	10ft/10ft	50	250 V	V	CS	840	BEESF-410	2	\$4,000	\$8,000	Ulrich 1984
T-405	Sterile Feed Tank	0.788	16ft/30ft	10	250 V	V	CS	183000	BEESF-410	1	\$60,000	\$60,000	Icarus 1985
T-410A/B	Cellulase Hold Tank		46ft/50ft	Ata	150 V	V	CS	214000	BEESF-510	1	\$67,000	\$67,000	Icarus 1985
FM-500A-AA	SSF Fermenter	0.948	46ft/50ft	Ata	150 V	V	CS	778000	BEESF-510	27	\$135,000	\$3,645,000	Icarus 1985
FM-501A	SSF Seed Fermenters (S.C.)	0.946	43ft/50ft	10	150 V	V	CS	451000	BEESF-510	1	\$122,000	\$122,000	Icarus 1985
FM-502A	SSF Seed Fermenters (S.C.)	0.945	19ft/30ft	10	150 V	V	CS	45100	BEESF-510	1	\$30,000	\$30,000	Icarus 1985
FM-503A	SSF Seed Fermenters (S.C.)	0.927	8ft/16ft	10	150 V	V	CS	4600	BEESF-510	1	\$10,000	\$10,000	Icarus 1985
FM-504A	SSF Seed Fermenters (S.C.)	0.948	4ft/7ft	10	150 V	V	CS	500	BEESF-510	1	\$3,400	\$3,400	Ulrich 1984
FM-505A	SSF Seed Fermenters (S.C.)	0.906	2ft/3ft	10	150 V	V	CS	70	BEESF-510	1	\$1,500	\$1,500	Ulrich 1984
FM-506A	SSF Seed Fermenters (S.C.)	0.777	1ft/1ft	10	150 V	V	CS	4	BEESF-510	1	\$500	\$500	Ulrich 1984
FM-507B	SSF Seed Fermenters (S.C.)	0.921	52ft/60ft	10	150 V	V	CS	952000	BEESF-510	1	\$269,100	\$269,100	Icarus 1985
FM-508B	SSF Seed Fermenters (S.C.)	0.952	20ft/38ft	10	150 V	V	CS	83200	BEESF-510	1	\$36,000	\$36,000	Icarus 1985
FM-509B	SSF Seed Fermenters (S.C.)	0.935	10ft/12ft	10	150 V	V	CS	9400	BEESF-510	1	\$14,000	\$14,000	Icarus 1985
FM-510B	SSF Seed Fermenters (S.C.)	0.941	5ft/7ft	10	150 V	V	CS	1000	BEESF-510	1	\$5,000	\$5,000	Icarus 1985
FM-511B	SSF Seed Fermenters (S.C.)	0.757	2ft/4ft	10	150 V	V	CS	94	BEESF-510	1	\$1,500	\$1,500	Ulrich 1984
FM-512B	SSF Seed Fermenters (S.C.)	0.707	1ft/2ft	10	150 V	V	CS	12	BEESF-510	1	\$900	\$900	Ulrich 1984
FM-513B	Seed Hold Tank (S.C.)	0.946	43ft/50ft	10	150 V	V	CS	451000	BEESF-510	1	\$122,000	\$122,000	Icarus 1985
T-501B	Seed Hold Tank (S.C.)	0.931	52ft/60ft	10	150 V	V	CS	764000	BEESF-510	1	\$190,000	\$190,000	Icarus 1985
T-507	Ethanol Vent Storage Tank	0.781	7ft/20ft	1.5	110 V	V	CS	4000	BEESF-510	1	\$15,000	\$15,000	Icarus 1985
T-508	Decasser Drum	0.482	7ft/20ft	15	300 V	V	CS	8200	BEESF-510	1	\$14,000	\$14,000	Icarus 1985
T-509	Beer Column Reflux Drum	0.555	4ft/10ft	50	250 H	H	CS	2500	BEESF-510	1	\$8,000	\$8,000	Icarus 1985
T-510	Rural Oil Decanter		4ft/10ft	25	150 H	H	CS	1300	BEESF-520	1	\$8,000	\$8,000	Icarus 1985
T-511	Rectification Column Reflux Dra	0.592	7ft/20ft	50	250 H	H	CS	2500	BEESF-520	1	\$8,000	\$8,000	Icarus 1985
T-512	Recycled Water Tank	0.727	4ft/14ft	50	250 V	V	CS	8800	BEESF-530	1	\$15,000	\$15,000	Icarus 1985
T-513	Ethanol Product Tank	0.817	7ft/20ft	Ata	150 V	V	CS	1023000	BEESF-710	2	\$287,300	\$574,600	Icarus 1985, Ulrich, 1982
T-514	Sulfuric Acid Storage Tank	0.856	52ft/52ft	Ata	150 V	V	CS	47000	BEESF-710	1	\$28,000	\$28,000	Icarus 1985
T-515	Fire Water Tank		16ft/40ft	250	150 H	H	CS	508000	BEESF-710	1	\$158,000	\$158,000	Icarus 1985
T-516	WHD Storage Tank	0.840	5ft/7ft	Ata	150 V	V	CS	59000	BEESF-710	2	\$85,000	\$170,000	Chemcost
T-517	Antifoam Storage Tank	0.731	15ft/15ft	Ata	150 V	V	CS	2500	BEESF-710	1	\$8,000	\$8,000	Icarus 1985
T-518	Diesel Fuel Tank		30ft/32ft	Ata	150 V	V	CS	21000	BEESF-710	1	\$48,000	\$48,000	Icarus 1985
T-519	Gasoline Storage Tank		20ft/15ft	Ata	150 V	V	CS	169000	BEESF-710	1	\$100,000	\$100,000	Icarus 1985, Peters 1980
T-520	Corn Steep Liquor Tank	0.935	15ft/20ft	15	650 V	V	CS	36000	BEESF-710	1	\$25,000	\$25,000	Icarus 1985
MS-605	Reactor Surge Drum		8ft/11ft	15	650 V	V	CS	16500	BEESF-810	1	\$32,000	\$32,000	Icarus 1985
MS-606	Offgas K.O. Suction Pot		8ft/11ft	15	650 V	V	CS	2800	BEESF-810	1	\$8,000	\$8,000	Icarus 1985
MS-607	Offgas Knock Out Drum		8ft/11ft	15	650 V	V	CS	2800	BEESF-810	1	\$8,000	\$8,000	Icarus 1985
MS-610	LP Vent Knock Out Drum		3ft/9ft	15	150 V	V	CS	470	BEESF-820	1	\$4,500	\$4,500	Ulrich 1984
T-802	Equalization Tank		82ft/40ft	Ata	70 V	V	CS	962000	BEESF-810	1	\$236,000	\$236,000	Icarus 1985
T-803	Anaerobic Reactor		30ft/50ft	15	650 V	V	CS	264000	BEESF-810	1	\$79,000	\$79,000	Icarus 1985
T-804	Bioreactor		74ft/30ft	Ata	70 V	V	Epoxy/CS	552000	BEESF-820	1	\$161,000	\$161,000	Icarus 1985
T-807	Blowdown Flash Drum		3ft/10ft	50	250 H	H	CS	350	BEESF-930	1	\$4,500	\$4,500	Ulrich 1984
MS-903	Hydrazine Drum		10ft/20ft	15	250 V	V	SS 316	250	BEESF-930	1	\$8,200	\$8,200	Icarus 1985
MS-904	Condensate Surge Drum		10ft/20ft	15	250 H	H	CS	11400	BEESF-930	1	\$24,000	\$24,000	Icarus 1985
MS-905	Plant Air Receiver		54ft/15ft	150	150 V	V	CS	2300	BEESF-940	1	\$27,500	\$27,500	Icarus 1985
MS-907	Instrument Air Receiver		50ft/40ft	Ata	150 V	V	CS	2300	BEESF-940	1	\$27,500	\$27,500	Icarus 1985
T-901	Process Water Tank		20ft/4ft	Ata	70 V	V	CS	587000	BEESF-920	1	\$154,000	\$154,000	Icarus 1985
T-905	Backwash Transfer Tank		2ft/10ft	250	500 V	V	CS/rubber	1600	BEESF-930	1	\$20,000	\$20,000	Icarus 1985
T-920	Condensate Collection Tank		10ft/17ft	Ata	200 V	V	CS	10000	BEESF-950	1	\$53,000	\$53,000	Icarus 1985
T-921	Sterile Water Tank		10ft/17ft	Ata	200 V	V	SS 304	10000	BEESF-960	1	\$14,000	\$14,000	Icarus 1985
T-922	Sterilization Tank		10ft/17ft	Ata	200 V	V	CS	10000	BEESF-950	1	\$53,000	\$53,000	Icarus 1985
T-923	Cleaning Tank		10ft/17ft	Ata	200 V	V	CS	10000	BEESF-950	1	\$14,000	\$14,000	Icarus 1985
T-924	Sterile Rinse Water Tank		10ft/17ft	Ata	200 V	V	CS	10000	BEESF-950	1	\$14,000	\$14,000	Icarus 1985

Total Equipment Cost \$10,449,700

SERI Proprietary Information  
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## **Appendix F**

### **Equipment List**

This section contains the equipment list for all equipment shown on the process flow diagrams. The list gives the equipment number, equipment name, specifications, cost, and source of the cost information.

Biomass To Ethanol  
Simultaneous Saccharification & Fermentation  
Project No. DOE/AFD/BESSF/02

EQUIPMENT LIST  
PUMPS

ITEM #	PUMP NAME	LB/HR	LB/CUFT	GPM	CAPACITY GPM	DELTA P psi	HYDRAUL. HP	PUMP EFFIC.	BRAKE HP	MOTOR HP/UNIT	MAT'L	TYPE	DRAW #	NO. REQ'D	PURCHASED COST	TOTAL PURCHASED COST	SOURCE
PP-101	Flume Pump	1000000	62.4	1998.0	2000	5	5.83	0.75	7.78	10	Cast Steel	Centrifugal	BESSF-110	1	\$4,300	\$4,300	Icarus 1985
PP-201A/S	Sulfuric Acid Pump	2878	112.4	3.2	4	32	0.07	0.4	0.19	1	SS	Recirculating	BESSF-210	2	\$3,200	\$6,400	Icarus 1987
PP-202A/S	Hydrolyzate Pump	942705	62.4	1883.5	2000	20	23.34	0.75	31.12	50	SS	Pos. Disp.	BESSF-210	2	\$67,000	\$134,000	Icarus 1985
PP-203A/S	Neutralized Hydrolyzate	944833	62.4	1887.8	2000	50	58.34	0.75	77.79	100	CS	Pos. Disp.	BESSF-210	2	\$63,000	\$126,000	Icarus 1985
PP-401A/S	Feed Pump	20125	62.4	40.2	525	20	6.13	0.7	8.75	15	CS	Pos. Disp.	BESSF-410	2	\$13,200	\$26,400	Icarus 1985
PP-403A/S	Fermenter Recycle Pump				750	50	21.88	0.7	31.26	40	CS	Centrifugal	BESSF-410	2	\$6,600	\$13,200	Icarus 1985
PP-411A/S	Prep Tank Transfer Pump	590	62.4	1.2	18	25	0.26	0.4	0.66	1	CS	Centrifugal	BESSF-410	2	\$2,100	\$4,200	Icarus 1985
PP-412A/S	Cellulase Feed Pump	29438	62.4	58.8	65	50	1.90	0.45	4.21	7.5	CS	Centrifugal	BESSF-410	2	\$3,300	\$6,600	Icarus 1985
PP-505A/S	Beer Transfer Pump	957399	62.4	1912.9	2000	30	35.01	0.75	46.67	60	CS	Centrifugal	BESSF-510	2	\$7,500	\$15,000	Icarus 1985
PP-509A/S	Ethanol Vent Pump	5259	49	13.4	15	60	0.53	0.4	1.31	2	CS	Centrifugal	BESSF-510	2	\$2,600	\$5,200	Icarus 1985
PP-601A/S	Beer Column Bottoms Pump	916653	62.4	1831.5	2000	60	70.01	0.75	93.35	125	CS	Centrifugal	BESSF-610	2	\$9,700	\$19,400	Icarus 1985
PP-603A/S	Beer Column Reflux Pump	141029	62.4	281.8	300	60	10.50	0.65	16.16	20	CS	Centrifugal	BESSF-610	2	\$5,000	\$10,000	Icarus 1985
PP-604A/S	Wash Return Pump				25	60	0.88	0.4	2.19	3	CS	Centrifugal	BESSF-620	2	\$3,000	\$6,000	Icarus 1985
PP-605A/S	Fusel Oil Pump	115	62.4	0.2	0.25	40	0.01	0.3	0.02	1	CS	Centrifugal	BESSF-620	2	\$2,100	\$4,200	Icarus 1985
PP-607A/S	Rectification Column Btms	60615	62.4	121.1	130	40	3.03	0.45	6.74	10	CS	Centrifugal	BESSF-620	2	\$4,300	\$8,600	Icarus 1985
PP-608A/S	Rectification Column Rflx	118014	49	300.3	320	60	11.20	0.65	17.23	25	CS	Centrifugal	BESSF-620	2	\$5,600	\$11,200	Icarus 1985
PP-631A/S	Recycled Water Pump	882129	62.4	1682.7	1700	30	29.75	0.75	39.67	50	CS	Centrifugal	BESSF-630	2	\$7,000	\$14,000	Icarus 1985
PP-632A/S	Sump Pump				100	20	1.17	0.45	2.59	5	CS	Centrifugal	BESSF-630	2	\$3,300	\$6,600	Icarus 1985
PP-701A/B/S	Ethanol Export Pump				525	30	9.19	0.7	13.13	20	CS	Centrifugal	BESSF-710	3	\$5,000	\$15,000	Icarus 1985
PP-703A/S	Sulfuric Acid Transfer				100	40	2.33	0.45	5.19	7.5	SS	Recirculating	BESSF-710	2	\$17,700	\$35,400	Chemcost
PP-704A/S	Fire Water Pump				200	100	35.01	0.7	50.01	60	CS	Centrifugal	BESSF-710	2	\$7,500	\$15,000	Icarus 1985
PP-706A/S	NH3 Transfer Pump				600	40	4.67	0.6	7.78	10	CS	Centrifugal	BESSF-710	2	\$4,300	\$8,600	Icarus 1985
PP-707A/S	Antifoam Transfer Pump				10	40	0.23	0.4	0.58	1	CS	Centrifugal	BESSF-710	2	\$2,100	\$4,200	Icarus 1985
PP-708A/S	Diesel Fuel Pump				25	50	0.73	0.4	1.82	3	CS	Centrifugal	BESSF-710	2	\$2,900	\$5,800	Icarus 1985
PP-710A/S	Gasoline Blending Pump				7.5	30	0.13	0.4	0.33	1	CS	Centrifugal	BESSF-710	2	\$2,100	\$4,200	Icarus 1985
PP-720A/S	Corn Steep Liquor Transfr				60	60	2.10	0.45	4.67	7.5	SS	Centrifugal	BESSF-710	2	\$3,500	\$7,000	Icarus 1985
PP-800A/S	Reactor Feed Pump				1100	50	32.09	0.75	42.78	60	CS	Centrifugal	BESSF-810	2	\$7,500	\$15,000	Icarus 1985
PP-809	Reactor Recycle Pump				1100	50	32.09	0.75	42.78	60	CS	Centrifugal	BESSF-810	1	\$7,500	\$7,500	Icarus 1985
PP-813A/S	Sludge Pump				255	20	3.09	0.65	4.75	7.5	CS	Pos. Displacement	BESSF-820	2	\$9,200	\$18,400	Icarus 1985
PP-816A/S	Final Effluent Pump				910	40	21.24	0.75	28.32	40	CS	Centrifugal	BESSF-820	2	\$6,500	\$13,000	Icarus 1985
PP-901A/S	Turbine Condensate Pump				200	40	4.67	0.6	7.78	10	CS	Centrifugal	BESSF-940	2	\$4,300	\$8,600	Icarus 1985
PP-902A/S	Process Water Transfer				2550	30	46.38	0.75	61.84	75	CS	Centrifugal	BESSF-920	2	\$8,800	\$17,600	Icarus 1985
PP-903A/S	Process Water Circulating				2550	60	92.77	0.75	123.69	150	CS	Centrifugal	BESSF-920	2	\$10,200	\$20,400	Icarus 1985
PP-904A/B	Backwash Feed Pump				6000	20	78.01	0.8	87.51	125	CS	Centrifugal	BESSF-920	2	\$9,700	\$19,400	Icarus 1985
PP-905A/B	Backwash Transfer Pump				50	20	0.58	0.4	1.46	2	CS	Centrifugal	BESSF-930	2	\$2,600	\$5,200	Icarus 1985
PP-906A/S	Blowdown Pump				50	25	0.73	0.4	1.82	3	CS	Centrifugal	BESSF-930	1	\$2,300	\$2,300	Icarus 1985
PP-907	Hydrazine Transfer Pump				5	20	0.06	0.35	0.17	1	SS	Centrifugal	BESSF-930	2	\$95,000	\$190,000	Icarus 1985
PP-908A/S	Boiler Feed Water Pump				1100	1250	882.22	0.75	1069.62	1200	SS	Centrifugal, staged	BESSF-930	2	\$95,000	\$190,000	Icarus 1985
PP-909A/S	Dearator Feed Pump				1100	20	12.84	0.75	17.11	25	CS	Centrifugal	BESSF-930	2	\$5,600	\$11,200	Icarus 1985
PP-910A/S	Condensate Pump				1100	30	19.25	0.75	25.67	40	CS	Centrifugal	BESSF-940	2	\$6,500	\$13,000	Icarus 1985
PP-912A-F/S	Cooling Water Pumps				9000	60	315.85	0.8	393.82	500	CS	Centrifugal	BESSF-940	7	\$15,600	\$109,200	Icarus 1985
PP-913A/S	Well Water Pumps				1800	25	25.25	0.75	35.01	50	CS	Centrifugal	BESSF-920	2	\$7,000	\$14,000	Icarus 1985
PP-953	Sterile Water Pump				10	30	0.18	0.4	0.44	1	CS	Centrifugal	BESSF-950	1	\$2,100	\$2,100	Icarus 1985
PP-950A/S	Supply Pump				20	50	0.58	0.4	1.46	2	SS	Centrifugal	BESSF-950	2	\$3,000	\$6,000	Icarus 1985
PP-955A/B/C/S	CIP/CS Sump Pump				20	30	0.35	0.4	0.88	2	SS	Centrifugal	BESSF-950	4	\$3,000	\$12,000	Icarus 1985

## PUMPS

Total Cost: \$1,005,100

Total HP Req'd: 5190

Biomass To Ethanol  
Simultaneous Saccharification & Fermentation  
Project No. DOE/AF/BEESF/02

EQUIPMENT LIST  
SOLIDS HANDLING

ITEM #	EQUIPMENT NAME	TYPE	DUTY/DESCRIPTION	HP/UNIT	MAT'L	DRAW #	NO. REQ'D	PURCHASED COST/UNIT	TOTAL PURCHASED COST	SOURCE
GS-103	Magnetic Chip Cleaner		remove down to .5" nuts	7	CS	BESSF-110	1	\$10,300	\$10,300	Icarus 1987
GA-101A/B/S	Front End Loaders	Diesel				BESSF-110	3	\$155,000	\$465,000	Icarus 1987
GS-101	Radial Stacking Conveyor	Paddle	1500 t/h wet wght, 120ft x 30in	15	CS	BESSF-110	1	\$124,500	\$124,500	Morbark 1983
GS-102	Belt Conveyor	Belt	240ft x 6.5ft wide, 200 t/h	10	CS	BESSF-110	1	\$191,700	\$191,700	Ulrich 1984
GS-104	Milled Chip Belt Conveyor	Belt	50ft X 6.5ft wide, 200 t/h	5	CS	BESSF-110	1	\$40,000	\$40,000	Ulrich 1984
8Y-101A/B/C/D	Wood Chip Unloader with Scale	23-ton/load	10 vans/hr/loader	35	CS	BESSF-110	4	\$33,400	\$157,600	Morbark 1983
GS-101A/B/C/D	Disk Refiner			2500	CS	BESSF-110	4	\$379,000	\$1,516,000	Sprout-Bauer Quote, 1990
GS-202	Screw Feeder	Auger	19500 cft/h	100	CS	BESSF-210	2	\$280,000	\$560,000	Black Clawson Quote 1990
GS-223	Lime Unloading Conveyor	Bucket	120 ft high, 100t/h	50	CS	BESSF-220	1	\$18,000	\$18,000	Peters et al. 80, Ulrich 84
GS-225	Lime Solids Feeder	Rotary Vlv	1.5 t/h	1	CS	BESSF-220	1	\$5,000	\$5,000	
MB-220	Lime Storage Bin		3500 cft		CS	BESSF-220	1	\$17,000	\$17,000	Icarus 1985
MF-224	Lime Unloading Pit		605 cft, 20ft X 50ft X 3ft		Concrete	BESSF-220	1	\$5,000	\$5,000	Means 1987
GS-611A/B	Sludge Screws	Screw, 18in	150ft long	10	CS	BESSF-630	2	\$20,500	\$41,000	Ulrich 1984
GS-001	Sludge Screws	Screw, 9in	100ft long	1.5	CS	BESSF-020	1	\$8,000	\$8,000	Ulrich 1984

SOLIDS HANDLING EQUIPMENT  
Total Cost: \$3,162,500  
Total HP Req'd: 10449.5

Biomass To Ethanol  
Simultaneous Saccharification & Fermentation  
EAS Project No. DOE/AFB/BESSF/02

EQUIPMENT LIST  
HEAT EXCHANGER

ITEM #	EQUIPMENT NAME	AREA SQ. FT.	TUBE		SHELL		TYPE	REMARKS	DRAW #	NO. REQ'D	PURCHASED COST/UNIT	TOTAL PURCHASED COST		SOURCE
			MAT'L/DP /TEMP deg F	psig	MAT'L/DP /TEMP deg F	psig								
TT-220	Feed Cooler	11362	CS/100		CS/100		Fixed Tube	Single pass	BESSF-220	1	\$22,600	\$22,600		Ulrich 1984
TT-312	Exhaust Condenser	1265	CS/100		CS/100		Vent Condenser		BESSF-310	1	\$14,100	\$14,100		Hall et al. 1982
TT-402	Water Cooler	80	CS/100		CS/100		Fixed Tube	Single pass	BESSF-410	1	\$3,000	\$3,000		Hall et al. 1988
TT-422	Fermenter Exhaust Cndnsr	685	CS/100		CS/100		Vent Condenser		BESSF-410	1	\$3,100	\$3,100		Hall et al. 1982
TT-525	Exhaust Condenser	1800	CS/100		CS/100		Vent Condenser		BESSF-510	2	\$17,600	\$35,200		Hall et al. 1982
TT-602	Degasser Drum Condenser	25	CS/100		CS/100		Vent Condenser		BESSF-510	1	\$3,400	\$3,400		Hall et al. 1982
TT-603	Beer Column Reboiler	1743	CS/100		CS/100		Reboiler		BESSF-510	1	\$58,800	\$58,800		Peters 1968, ICARUS
TT-605	Beer Column Condenser	1400	CS/100		CS/100		Fixed tube		BESSF-510	2	\$15,500	\$31,200		ICARUS 1985
TT-606	Beer Column Vent Condensr	36	CS/100		CS/100		Vent Condenser		BESSF-510	1	\$3,400	\$3,400		Hall et al. 1982
TT-607	Fusel Oil Cooler	4	CS/100		CS/100		Double pipe		BESSF-620	1	\$400	\$400		
TT-609	Rectification Clm Reboiler	980	CS/100		CS/100		Reboiler		BESSF-620	1	\$20,800	\$20,800		Icarus 1985
TT-610	Rectification Clm Cndsr	1785	CS/100		CS/100		Fixed tube		BESSF-620	1	\$20,800	\$20,800		Hall et al. 1988
TT-611	Rectfctn Clm Vent Cndsr	43	CS/100		CS/100		Vent Condenser		BESSF-620	1	\$3,400	\$3,400		Hall et al. 1982
TT-613	Feed Preheater	750	CS/100		CS/100		Fixed tube		BESSF-610	1	\$9,600	\$9,600		Hall et al. 1988, Pe
TT-615	Feed Cross Exchanger	1331	CS/100		CS/100		Floating head		BESSF-610	1	\$35,400	\$35,400		Hall et al. 1988, Pe
TT-801	Offgas Cooler	500	CS/100		CS/100		Fixed tube		BESSF-810	1	\$8,000	\$8,000		Hall et al. 1988
TT-802	Feed Cooler	1600	CS/100		CS/100		Fixed tube	4 shell-8 tube pass	BESSF-810	5	\$16,600	\$83,000		Hall et al. 1988
TT-953	Water Sterilizer	1.25	CS/100		CS/100		Double pipe		BESSF-950	1	\$200	\$200		
	Xylose Fer. Coils	2257	CS/100				Coils	4.0 in OD, Sch 10	BESSF-310	8	\$2,900.00	\$23,200		Vendor Quote
	Cellulase Fer. Coils	8627	CS/100				Coils	6.0 in OD, Sch 10	BESSF-410	3	\$13,900	\$41,700		Vendor Quote
	SSF Fer. Coils	692	CS/100				Coils	2.5 in OD, Sch 10	BESSF-510	27	\$1,000	\$27,300		Vendor Quote

Total Equipment Cost \$448,300

Biomass To Ethanol  
Simultaneous Saccharification & Fermentation  
Project No. DOE/AF/BESEF/02

EQUIPMENT LIST  
MISCELLANEOUS

ITEM #	EQUIPMENT NAME	DUTY/DESCRIPTION	HP/Unit	MAT'L	REMARKS	DRAW #	NO. REQ'D	PURCHASED COST/UNIT	TOTAL PURCHASED COST	SOURCE
GA-201	Line Mixer			Hastalloy		BESEF-210	1	\$2,100	\$2,100	Icarus 1987
GA-203	Blowdown Tank Agitator	Single Impeller	25	SS 304		BESEF-210	1	\$18,300	\$18,300	Chemcost
GA-213	Neutralization Tank Agitator	Single Impeller	50	SS 304		BESEF-220	1	\$29,600	\$29,600	Chemcost
GC-227	Lime Dust Cyclone	250 ASCFM, 15 lb/h solids		CS		BESEF-220	1	\$600	\$700	Chemcost
GF-201	Desiccant Air Filter	0.5 cfm		Silica		BESEF-210	1	\$1,800	\$1,800	
MR-201	Impregator with Rotary Valve	2564 cu ft, 20 HP drive, 20 HP rtry vlv	40	C-20 Cb3	Continuous pulp digester	BESEF-210	2	\$1,630,000	\$3,660,000	Black Clawson 1990
MR-202	Prehydrolysis Reactor	2564 cu ft, 20 HP drive, 20 HP rtry vlv	40	C-20 Cb3	Continuous pulp digester	BESEF-210	2	\$1,630,000	\$3,660,000	Black Clawson 1990
GA-301	Seed Hold Tank Agitator	Single Impeller/ 40 hp	20	CS		BESEF-310	1	\$16,300	\$16,300	Chemcost
GA-303A-H	Xylose Fermenter Agitator	Single Impeller	75	CS		BESEF-310	8	\$28,300	\$231,200	Chemcost
GA-305	First Seed Vessel Agitator	Single Impeller	150	CS		BESEF-310	1	\$55,200	\$55,200	Chemcost
GA-306	Second Seed Vessel Agitator	Single Impeller	40	CS		BESEF-310	1	\$16,300	\$16,300	Chemcost
GA-307	Third Seed Vessel Agitator	Single Impeller	5	CS		BESEF-310	1	\$6,800	\$6,800	Chemcost
GA-308	Fourth Seed Vessel Agitator	Single Impeller	0.5	CS		BESEF-310	1	\$2,000	\$2,000	Chemcost
GA-400	Prep Tank Agitator	Single Impeller/ 10 hp	0.6	CS		BESEF-410	1	\$7,600	\$7,600	Chemcost
GA-401A/B/C	Fermenter Agitator	Single Impeller	200	CS		BESEF-410	3	\$75,300	\$225,900	Chemcost
GA-405	Feed Tank Agitator	Single Impeller/ 300 hp	150	CS		BESEF-410	1	\$127,000	\$127,000	Chemcost
GA-410	Hold Tank Agitator	Single Impeller/ 200 hp	100	CS		BESEF-410	1	\$75,300	\$75,300	Chemcost
GA-411	First Seed Vessel Agitator	Single Impeller	40	CS		BESEF-410	2	\$16,300	\$33,600	Chemcost
GA-412	Second Seed Vessel Agitator	Single Impeller	1.5	CS		BESEF-410	2	\$3,000	\$6,000	Chemcost
GA-413	Third Seed Vessel Agitator	Single Impeller	0.25	CS		BESEF-410	2	\$1,500	\$3,000	Chemcost
GA-500A-AA	SSF Fermenter Agitator	Single Impeller	80	CS		BESEF-510	27	\$29,300	\$807,300	Chemcost
GA-501A	Seed Hold Tank Agitator (S.c.)	Single Impeller/ 50 hp	25	CS		BESEF-510	1	\$20,100	\$20,100	Chemcost
GA-501B	Seed Hold Tank Agitator (B.c.)	Single Impeller/ 75 hp	35	CS		BESEF-510	1	\$28,900	\$28,900	Chemcost
GA-510A	First Seed Vessel Agitator (S.c.)	Single Impeller	250	CS		BESEF-510	1	\$97,200	\$97,200	Chemcost
GA-511A	Second Seed Vessel Agitator (S.c.)	Single Impeller	50	CS		BESEF-510	1	\$20,100	\$20,100	Chemcost
GA-512A	Third Seed Vessel Agitator (S.c.)	Single Impeller	7.5	CS		BESEF-510	1	\$6,800	\$6,800	Chemcost
GA-513A	Fourth Seed Vessel Agitator (S.c.)	Single Impeller	0.75	CS		BESEF-510	1	\$2,500	\$2,500	Chemcost
GA-510B	First Seed Vessel Agitator (B.c.)	Single Impeller	400	CS		BESEF-510	1	\$172,600	\$172,600	Chemcost
GA-511B	Second Seed Vessel Agitator (B.c.)	Single Impeller	75	CS		BESEF-510	1	\$28,900	\$28,900	Chemcost
GA-512B	Third Seed Vessel Agitator (B.c.)	Single Impeller	10	CS		BESEF-510	1	\$7,500	\$7,500	Chemcost
GA-513B	Fourth Seed Vessel Agitator (B.c.)	Single Impeller	1	CS		BESEF-510	1	\$2,700	\$2,700	
GC-609A/B/C	Centrifuge	Solid Bowl	200	CS	Assumed 75% SS cost	BESEF-630	3	\$225,000	\$675,000	Badger 1984
GF-783	Desiccant Air Filter	1500 cfm		Silica		BESEF-710	1	\$38,000	\$38,000	Icarus 1985
GC-801	Sludge Centrifuge	Solid Bowl	40	CS	Assumed 75% SS cost	BESEF-820	1	\$128,000	\$128,000	Badger 1984
GO-806	Offgas Burner			CS		BESEF-810	1	\$20,000	\$20,000	
GV-807	Biotreater Agitators			Polyethylene		BESEF-820	1	\$10,000	\$10,000	
GV-808	Secondary Clarifier	Center feed, 100 ft diameter	3	CS		BESEF-820	1	\$260,000	\$260,000	Ulrich 1984
PB-810	Offgas Blower	2530 cfm, 20 psig discharge	60	CS		BESEF-810	1	\$74,300	\$74,300	Peters and Timmerhaus 1980
PB-812A/S	Aeration Blowers	1300 cfm, 25 psig discharge	60	CS		BESEF-820	2	\$59,400	\$118,800	Peters and Timmerhaus 1980
PB-817A/S	LP Vent Blower	2400 cfm, 20 psig discharge	60	CS		BESEF-830	2	\$73,000	\$146,000	Peters and Timmerhaus 1980
GA-900	Sterilization Tank Agitator	Single Impeller	10	SS 304		BESEF-900	1	\$10,000	\$10,000	Chemcost
GA-901	Cleaning Tank Agitator	Single Impeller	10	SS 304		BESEF-900	1	\$10,000	\$10,000	Chemcost
GF-901	Sand Filter	34ft dia. X 8ft high		CS		BESEF-920	1	\$39,600	\$39,600	Icarus 1985
GT-912	Cooling Tower System	54000 gpm	110	CS	0.6 scaling factor	BESEF-940	1	\$751,400	\$751,400	Badger 1984
GU-903A/B	Demineralizers	200 gpm		CS	0.6 scaling factor	BESEF-930	2	\$308,200	\$616,400	Badger 1984
GU-904A/S	Condensate Polisher	1400 gpm		SS		BESEF-930	2	\$100,000	\$200,000	
GU-907	Hydrazine Addition Package	150 gal tank, 2 pumps, 1 agitator	1	SS		BESEF-930	1	\$15,000	\$15,000	
GU-908	Ammonia Addition Package	150 gal tank, 2 pumps, 1 agitator	1	SS		BESEF-930	1	\$15,000	\$15,000	
GU-909	Phosphate Addition Package	150 gal tank, 2 pumps, 1 agitator	1	SS		BESEF-930	1	\$15,000	\$15,000	
GV-906	Dearator	1700 gpm, 17000 gal		CS shell/ SS internals		BESEF-930	1	\$133,000	\$133,000	Badger 1984
GV-910	Instrument Air Dryer	600 scfm/ desiccant		CS		BESEF-940	1	\$23,100	\$23,100	Icarus 1985
GZ-911	Turbo Generator	25 MW				BESEF-940	1	\$6,500,000	\$6,500,000	ABB Quote 1990
HB-901A	Steam Boiler	1100 psia, 450000lb/h, 300F superheat			\$19,000.00 Installed	BESEF-910	1	\$0	\$0	ABB Quote 1990
PC-911	Air Compressor	1000 scfm	250	CS	0.7 scaling factor	BESEF-940	1	\$58,300	\$58,300	A.D. Little 1984
PK-950A/B/S	Air Compressor Package	28000 scfm	2500	CS	0.7 scaling factor	BESEF-950	3	\$600,000	\$1,800,000	A.D. Little 1984
PK-951	Chilled Water Package	3300 gpm, 50 F	2360	CS		BESEF-950	1	\$600,000	\$600,000	

Total Equipment Cost \$21,634,300  
Total HP Required 16785

Biomass To Ethanol  
Simultaneous Saccharification & Fermentation  
Project No. DOE/AFB/BESSF/02

EQUIPMENT LIST  
TOWERS

ITEM #	EQUIPMENT NAME	DIA/HTGT (feet)	NO. TRAYS	PRESS. psi	TEMP. F	MAT'L	REMARKS	DRW #	NO. REQ'D	PURCHASED COST/UNIT	TOTAL PURCHASED COST	SOURCE
AS-601	Beer Column	17.0/32	16	15	300	CS		BESSF-610	1	\$205,100	\$205,100	Icarus 1985
AS-602	Rectification Column	13.5/48	24	15	300	CS		BESSF-620	1	\$192,500	\$192,500	Icarus 1985
Total Equip. Cost: \$397,600												